Global Optimization and Thermodynamic Analysis of the Heat Exchanger Network in Process Industries

BY

AMIR ABDELRAZIG MERGHANI MOHAMMED

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This thesis, written by **Amir Abdelrazig Merghani Mohammed** under the direction of his thesis advisor and approved by his thesis committee, has been presented and accepted by the Dean of Graduate Studies, in partial fulfillment of the requirements for the degree of **DOCTOR OF PHILOSOPHY IN CHEMICAL ENGINEERING.**

Ealmutairi

Dr. Eid M. Al-Mutairi
(Advisor)

Dr. Mohammed S. Ba-shammakh
Department Chairman

Dr. Housam Binous
(Member)

Dr. Salam A. Zummo
Dean of Graduate Studies

DEANSHIP OF GRADUATE STUDIES

Dr. Mohammed S. Ba-shammakh
(Member)

14/9/17

Date

Mozzhar Hossain

Dr. Mohammad M.Hossain
(Member)

C. Basha

Dr. Basheer Chanbasha
(Member)

Dedication

This work is dedicated to my beloved parents, wife, brothers, sisters, and friends for support and prayers

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I express my gratitude to Almighty ALLAH, the most Beneficent and the most Merciful. He gave me healthy, support and mental strength to complete this work. Great respect to Holy Prophet Muhammad peace be upon him whose advises and orders to learn knowledge in all circumstances.

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LIST OF ABBREVIATIONS

Symbols	Meaning (Units)
A	Area of heat exchanger
a & b	Denote the process stream and utility stream in HENs
CP	Heat capacity flow rate (MW/K)
Cp	Specific heat capacity (Kj/Kg.K)
F*Cp	Heat capacity flow rate (MW/K)
ΔH	Exchanger duty (MW)
HEN	Heat exchanger network
HENs	Heat exchanger networks
HEs	Heat exchangers
MM	Millions
Q	Duty
R	Interest rate (%)
T^T	Outlet temperature of stream (K)
T^S	Inlet temperature of stream (K)
U	Overall heat transfer coefficient (MW/m2 K)
N_{HX}	Total number of heat exchangers
C_{CU}	Cost per unit of cold utility
C_{HU}	Cost per unit of hot utility
C_k	Exponent for area cost

Indices

i hot process or utility stream

j cold process or utility stream

K index for a heat exchanger, 1 ……. N_{HX}

Binary variables

ψ_{ijk} indicating the existing of match ij at the heat exchanger, k, in optimal network

$\psi_{i,CU}$ indicating the existing of the match between hot stream i and cold utility

$\psi_{\mathrm{HU,j}}$ indicating the existing of the match between hot utility and cold stream j

Variable

$Q_{\mathrm{i,CU}}$ heat exchanged between hot stream i and cold utility

$Q_{\mathrm{HU,j}}$ heat exchanged between hot utility and cold stream j

ABSTRACT

Full Name : Amir Abdelrazig Merghani Mohammed

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Energy conservation and reducing its environmental impact are crucial for process design and synthesis nowadays. In this study, an energy optimization analysis, along with targeting, and retrofitting of an existing complex industrial petrochemical plant were conducted. Stochastic optimization methods were applied to achieve a highly sustainable energy integration system. Two techniques, (1) Simulated Annealing (SA) and (2) fixed structure, were used to retrofit the existing HEN and to lower the Total Annual Cost (TAC). In SA the HEN is flexible and a new heat exchanger can be added or removed. While in the fixed structure technique, the system allows for a very limited space for changing the number of heat exchangers in the network. The existing network of PDH is composed of 13 hot and 12 cold streams. The hot and cold utility demands of the existing network were 72.54 [MW] and 128.53 [MW], respectively. A revamp design was constructed based on the SA method and the fixed structure techniques. The results of the SA optimization were superior when compared to the fixed structure counterparts. The suggested new design has a savings of 20.65 million US $/year (savings up to 14.64%). Moreover, heat exchanger network of a fractionation unit is composed of 6 hot and 7 cold streams. The hot and cold utility demands of the existing network were 57.42 [MW] and 74.68 [MW], respectively. A revamp design was constructed based on the Simulated Annealing method (SA) and the fixed structure techniques. The results of the fixed structure were superior to their SA optimization counterparts. The suggested modern design has a saving of 5.58 million US $/year (save up to 21.45%). Optimization scenarios are presented by putting some constraints on the HEN. An algorithm for utility exchanger modification was generated using the Aspen Energy Analyzer's Heat Exchanger Network Grid Diagram software that resulted in the optimized design after network evaluation. The design was optimized depending upon the capital cost and energy recovery. The design options were compared and the best retrofit option for heat exchange networks (HENs) was selected. The best design that was suggested was A-Design1- 4U that has a reduced payback period of 4.52 years and offers maximum energy savings

ملخص الرسالة

الاسم الكامل: عامر عبدالرازق ميرغنى محمد

عنوان الرسالة: التحسين العالمي والتحليل الديناميكى الحرارى لشبكه المبادلات الحرايه في العمليات الصناعية

التخصص: الهندسة الكيميائية

تاريخ الدرجة العلمية: مايو 2017

قد اصبح من المهم الحفاظ على الطاقه والحد من تاثيرها البيئى اثناء تصميم العمليات وتوليفها فى الوقت الحاضر. في هذه الدراسة، تم إجراء التحليل الأمثل للطاقة، والهدف منه، وإعادة تجهيز مصنع البتروكيماويات الصناعية المعقدة. وعليه تم تطبيق أساليب التحسين العشوائي لتحقيق نظام متكامل للطاقة المستدامة الطويله الامد. تم استخدام تقنيتين، (1) تركيب المحاكاة القوى (SA) و (2) وتقنية البنية الثابتة ، استعملت في إعادة تأهيل شبكه المبادلات الحراريه (HENs) الحاليه وخفض التكلفة السنوية الإجمالية (TAC). فى تقنيه المحاكاة القويه (SA) هنالك مرونه فى شبكه المبادلات الحراريه (HEN) فيمكن إضافة مبادل حراري جديد أو إزالته ،بينما فى تقنيه البنيه الثابثه فان النظام له مساحة محدودة جدا لتغيير عدد المبادلات الحرارية في الشبكة. وتتألف الشبكة الحاليه من 13 تيارات ساخنة و 12 تيارات باردة. وكانت متطلبات المرافق الساخنة والباردة للشبكة القائمة 72.54 [ميغاواط] و 128.53 [ميغاواط] على التوالي. تم اعاده التصميم على أساس طريقة المحاكاة القويه (SA) وتقنيات الهيكل الثابت. وكانت نتائج التحسين الناتجة من (SA) متفوقة بالمقارنة مع نظيرتها البنية الثابتة. التصميم الجديد المقترح لديه توفير 20.65 مليون دولار أمريكي / للسنة (حفظ ما يصل الى 14.64٪).

واضافة على ذلك، تتكون شبكة المبادلات الحرارية لوحدة تجزئة من 6 تيارات ساخنة و 7 تيارات باردة. وكانت متطلبات المرافق الساخنة والباردة للشبكة الحالية 57.42 [ميقاواط] و 74.68 [ميقاواط]، على التوالي. تم اقتراح تصميم جديد على أساس طريقة محاكاة القويه (SA) وتقنيات الهيكل الثابت. وكانت نتائج الهيكل الثابت متفوقة على نظيرتها المحاكاة القوية (SA). التصميم الحديث المقترح قد وفر 5.58 مليون دولار أمريكي / للسنة (حفظ ما يصل الى 21.45٪). يتم عرض سيناريوهات التحسين من خلال وضع بعض القيود لتحسين شبكة المبادات الحرارية. تم إنشاء خوارزمية لتعديل بطريقة مبادل المرافق باستخدام برنامج الاسبن (Aspen Energy Analyzer's) لشبكة المبادلات الحرارية بعد التحسين الاول للشبكه الذى حدث لها وتقيم الشبكة. وقد تم تحسين التصميم اعتمادا على التكلفة الرأسمالية واسترداد الطاقة. تمت مقارنة الخيارات المصممة واختيار الأفضل للشبكة . وأفضل تصميم تم اقتراحه هو A-Design1- 4U الذى لديه فترة استرداد اقل وهى 4.52 سنة وتوفر أقصى قدر من الطاقة مقارنه بالتصاميم الاخرى .

CHAPTER 1
INTRODUCTION

Petrochemical industries are the main part of an economy in both developed and developing countries alike. Also, continuous efforts are made to maximize the heat recovery while saving energy with minimum harm to the environment [1]. To achieve this, energy optimization techniques can be used to determine the thermodynamic properties [2]. The occurrence of energy crises has motivated both industrialists and academia to begin striving for effective research and development in the field of the economic utilization of energy in process industries[3]. This situation, coupled with a growing need for a pollution-free environment, has resulted in research efforts in energy optimization and best process integration for the optimum use of energy. This has been an encouraging trend all over the world[4]. Tools, such as pinch analysis, have been widely applied to analyze process plants, including petrochemical plants. However, some works on pinch analysis have attempted to generate an optimal heat exchanger network for systems and to develop mathematical models to minimize Total Annual Cost (TAC) together with certain assumptions which relied on constant process parameters [5]–[8]. The assumptions regarding constant process stream properties result in conclusions that are far away from industrial realities. However, increasing attention has been focused on the application of pinch concepts in the process to reduce the total annual operating cost [9]. In the approach used by Floudas and Grossmann (1987) to address this problem, various models were adopted[10]. For instance, Linear Programming (LP) was used to find out the lowest utility cost and the location of the pinch points for each of the chosen parameters. Also, Mixed Integer Linear Programming (MILP) was used for the minimum unit determination. However, it should be noted that the decomposition of the Heat Exchanger Network (HEN) problem into separate targeting procedures, i.e. utility, energy, and the area, does not guarantee total cost minimization. In this research, a Mixed Integer Nonlinear Programming (MINLP) optimization model is used, which is qualified to capture the common features of the HEN and guarantee minimization of the total/overall cost[11]. The MINLP follows the mathematical formulations of Yee and Grossmann (1990), with an improvement in the model by considering the variations in the process steam properties[12].

Generally, the purpose of process optimization is to increase the plant throughput by operating the plant at or near the optimum conditions. By combining all these concepts, along with their usefulness into an optimal and cost effective operation of a petrochemical plant, this work aims to bridge the existing gaps in the previous studies by conducting analysis of petrochemical units, establishing generic optimum conditions based on varying process properties rather than the assumption of constant scenarios, and quantifying the extent of the process streams' inefficiencies. Strong competition in the ever dynamic international market and stringent product requirements are the driving force compelling the process industries to look keenly into plant operation in a profitable manner. To avert potential dwindling profit margins due to high competition, and to maintain market share in the international arena, the need for real-time or on-line optimization of the entire plant is equally essential and hence covered in this study.

According to the current report by the Organization of the Petroleum Exporting Countries (OPEC) (2012), Saudi Arabia possesses 18 % of the world's proven petroleum reserves and is ranked as the largest exporter of petroleum. For this country to maintain its exalted position in OPEC, the need for conducting research that can improve the performance of the existing facilities is justified and should be in high demand. Globally, the general energy supply and environmental situation require improved management for the utilization of energy resources. Improved petrochemical plant efficiency for optimum production at minimum energy consumption is capable of reducing carbon dioxide emissions associated with fossil-fuel combustion. In the global sense, this research work also includes the environmental protection aspect which is currently a hot topic of discussion worldwide.

This study is limited to the development of an improved mathematical model for the optimum performance of the petrochemical process at minimum TAC. The performance of the improved mathematical models developed was quantified by comparing their results with the existing models. The test of the improved and existing models was done using examples from literature; the improved model was thereafter applied to case studies. The specific case study does not intend to approach the design of the petrochemical industry from scratch; rather, it focuses on the analysis of the functional petrochemical operations in the Kingdom of Saudi Arabia. The process data for the analyses was obtained for the Kingdom through exploration of the existing linkages between KFUPM and the petrochemical industries in Saudi Arabia. This work further strives to specify some basic thermodynamic parameters for the optimum performance of the plant that were missing in the previous literature, in addition to the detailed, energy, and economic analyses of the plant.

This study also provides information on the methods for the reduction of energy consumption and emissions. The findings of this study will help in effective energy savings for the entire plant, including processed-bottlenecking, investment cost reduction, process modification and total site planning.

CHAPTER 2
LITERATURE REVIEW

The main purpose of this review is to present information on the state of the art scientific literature on process integration and real-time optimization as they relate to the proposed research. This review also intends to identify gaps in knowledge in the previous research that the current research seeks to bridge.

2.1 Pinch Technique

Worrell and Galitsky suggested that, for any plant that has various heating and cooling streams, the pinch technique can be used to increase the overall efficiency [13]. The pinch method was first presented by Linnhoff and Vredeveld (1984), and it uses thermodynamic calculations as well as optimization algorithms to maximize the heat recovery for Heat Exchanger Networks (HENs) [14]. The pinch method is based on the practical concept of the first and second laws of thermodynamics. It is known that, in general, whatever the objective may be in a design (e.g. capital cost reduction, improving arability etc.), the engineer starts from the material balance which is derived from the concept of the first law of thermodynamics [15].

The key concept of process integration is the heat transfer from the hot and cold streams within the network before using the hot and cold utilities. It is a very effective technique to reduce energy consumption and is also environmentally friendly [16], [17]. The optimum operation values can change from time to time, so it is preferable to recalculate the pinch analysis to achieve better results for the petrochemical plant. Strang and Linnhoff-Popien (2004) claimed that using the pinch analysis for any industrial process can save utility energy by up to 30%, accordingly, we can save up to 15% from the economic point of view[18]. Also, Al-Mutairi has used the pinch analysis technique to modify the existing heat exchanger network in a fluid catalytic cracking unit, by using the optimum ΔT_{min}[19]. The proposed design has not only increased the heat recovery of the unit but also reduced the energy cost. Nowadays, there is an increasing focus on the application of pinch concepts in the formulation of mathematical models in order to reduce the total annual operating cost [1], [20].

The implementation of pinch analysis starts with calculating the minimum energy utilization for the whole process, then there will be two networks (below and above pinch point). To find the minimum units for the HENs each network is treated separately using the heuristics method.

2.2 Heat Exchanger Network Synthesis (HENS) mathematical model

The basis for the heat exchanger network mathematical model is drawn from recognition of the existence of an analogy between the HENS and the transshipment model [21]. In this approach, a chemical process comprises different process streams/utilities, including some which need to be heated while others need to be cooled. Heat available in the hot streams/utilities acts as a commodity to be shipped, hot streams or utilities operate as sources, cold streams or utilities operate as destinations, and temperature intervals (TI) act as intermediate warehouses. This analogy is depicted in Figure 2.1.

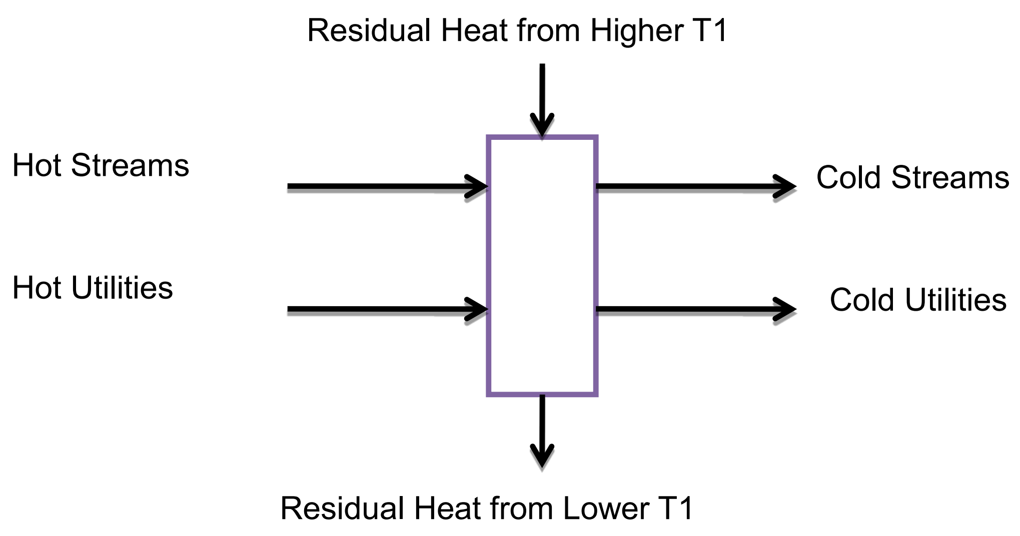

Figure 2.1: Analogy between HENS and transshipment model

From the figure, it can be observed that the hot streams can transfer heat to the cold streams and utilities. There also exists the possibility of heat transfer from the hot utilities to the cold streams. Further, the hot streams can cascade their residual heat to a lower temperature interval (TI). Several methods and approaches either sequential or simultaneous have been developed to synthesize a heat exchanger network that yields a commendable trade-off between capital and cost of operation. Yee and Grossmann (1990) presented an (MINLP) model that could create networks with optimized heat exchanger area and cost[12]. The model procedure does not depend on the fixed temperature or pinch point estimations that decompose the whole system to sub-networks, as in the case of the pinch design method. The model relied on the stage-wise technique in which heat exchange could appear within each stage for the available hot and cold lines and potential hot and cold matches. Galli and Cerdá (1998) proposed an MILP framework with a new decomposition strategy [22]. In a similar vein, Briones and Kokossis (1999) presented a synthesis method that combines thermodynamics and mathematical programming [20]. Ponce-Ortega et al., (2008), introduced an MINLP model for a heat exchanger network that has different phases within the streams[23].

Huang et al., 2012, synthesized HEN using a stage wise technique[24]. Moreover, they proposed an approach that can easily determine the log-mean temperature difference (LMTD). The interval based MINLP model was developed by Isafiade and Fraser (2010) for operations that have different intervals[8]. It was discovered that all the approaches cited above were all based on the assumption of fixed operating parameters such as the heat capacity flow rate, and known supply and target temperatures [16]. Another important factor to be considered is that these approaches were applied to literature problems involving only a few hot and cold streams. The proposed work is aimed at synthesizing a HEN in the operating petrochemical industry units. The generated network will be highly robust to handle unpredictable parameter changes that can arise from sessional changes or operational variations from the start, middle, and end of the operation. For a general consideration of the regions where there are no sharp sessional changes for winter and summer, these periodical changes will be regarded as the start-up, the middle of an operation and the shut-down morning [17].

2.3 Objectives

The main objective of this research is to enhance the design of the heat exchanger networks for petrochemical industries.

The specific objectives are to:

i. Apply pinch technology to formulate robust mathematical models in order to revamp the existing heat exchanger networks.

ii. Use the process simulation software i-heat to study the sensitivity analysis of the heat exchanger networks via superstructure and Pinch configuration techniques etc.

iii. Use GAMS/ i-heat to solve the formulated MINLP model to optimize the heat exchanger networks in terms of minimizing the TAC.

iv. Conduct an extended analysis of the plant beyond an analysis of the process components, by including the contribution of the process streams and the environmental influence on process performance.

v. Optimize the design after network evaluation using the Aspen Analyzer software and the pinch technique by modifying the utility heat exchanger method to trade off capital cost with energy recovery.

2.4 Scope of the this work

The study is broadly divided into two phases as described below

2.4.1 Model formulation

This first phase focuses primarily on an extensive review of the relevant literature materials on heat exchanger network synthesis and modeling of the interaction between the heat exchanger network and the process components. Mostly, the synthesis of existing models on heat exchanger networks was based on constant process parameters [5], [8], [15]. The mathematical relevance of these models was assessed by using GAMS/i-heat.

2.4.2 Application to petrochemical industry cases

At this stage, the design and operating data which were acquired from the existing process were used to develop a steady state simulation model of the process using i-Heat process simulation software. Heat and mass balances data generated at this phase was used for the optimization of the plant using GAMS. In addition, this involves the synthesis of a multi-period, flexible heat exchanger network for minimum TAC, for varying the process stream temperature, heat exchanger duty and heat exchanger area. Environmental influence on process operations could result in fluctuations in the process parameters. The detailed analyses which were performed are essential to quantify the effect of these process stream property fluctuations. These analyses lead to the generation of thermodynamic data that could stand as a reference for plant designers towards the design of process equipment suitable for handling such fluctuations for global optimum operation

CHAPTER 3

METHODOLOGY

In this work, an improved heat exchanger networks (HENs) for optimum performance of the Propane Dehydrogenation (PDH) process at a minimum total annual operating cost is developed. Two techniques were adopted to optimize and retrofit the HEN, (1) the Simulated Annealing (SA) and (2) fixed structure technique using MINLP formulation. In the simulated annealing technique, the HEN is flexible and a new heat exchanger can be added or removed. However, in the fixed structure technique, the system allows a very limited space for changing the number of heat exchangers in the network. The results of optimization can be applied to the PDH process by selecting a feasible revamp scheme. The results of our improved HENs will be compared to those of the existing HEN.

Here, the design and operating data, which were acquired from an existing process, were used to develop the steady state simulation of the process using i-heat process simulation software. Heat and mass balance data generated during the above phase were then used in order to optimize the PDH plan. The methodology adopted in this study involves revamping optimization, mathematical model's formulation, algorithm development, simulation and application of models to solve optimization problems in petrochemical operations.

3.1 Mathematical Model Formulation

Mathematical equations were formulated to model the heat exchange process within the process streams (heat exchanger network), fluid flow and in the process equipment e.g. fractionation unit. The existing models have been improved, whereas the process stream properties were considered constant. The mathematical formulation of an optimization model comprises the following steps:[16]

a) Determination of the objective function

In order to determine the objective function, the following points are generally taken into consideration:

- Determine the amount to be optimized.
- Classify the decision (optimization) variables that are required to define the objective function.
- Describe the objective function mathematically in terms of the optimization variables.

b) Development of a game plan

A game plan is developed to tackle the problem at hand and generally includes some imperative considerations, such as:

- Determining a convenient technique to address the problem.
- Specifying the reasonable aspiration levels according to decision- maker's point of view.
- The key concept necessary to transfer our opinions and method into a working formulation.

c) Model constraints

The following steps are generally considered when developing model constraints:

- Transform the assigned approach into mathematical expressions.
- Create a search space that can establish solution alternatives.

- Identify all straightforward relations, and restrictions, as well as limitations to be expressed mathematically as equality and inequality constraints. These are required for the approach description.

Associate masterful constraints such as non-negativity or integer requirement on the input parameters.

d) Effectively formulated model

To come up with an effective model, the decision-maker should consider the following:

- Avoid highly non-linear constraints or terms that may end up with challenges in solving the developed model.
- Improve the model formulation to be understandable in order to reveal significant information.

3.2 Algorithm Development

Effective programming requires the development of efficient algorithms that spell out the series of steps for the computer codes in the ways that are easily understood. In this the study, the mathematical equations formulated were subsequently written to execute the algorithm to generate the results.

The objective:

The (TAC) is the objective function which contains the hot utility cost, cold utility cost, exchanger fixed duty and the area cost which is defined as follows, by a modified version of the equation used by Al-Mutairi and Odejobi [1].

$$\begin{aligned}&Min\ \mathrm{TAC} = C_{CU}\sum_{i=1}^{N_{hot}}\sum_{k=1}^{N_{HX}} U_{i,CU,k} A_k LMTD_{i,CU,k}\psi_{i,CU,k} + C_{HU}\sum_{j=1}^{N_{cold}}\sum_{k=1}^{N_{HX}} U_{\mathrm{HU,j},k} A_k LMTD_{\mathrm{HU,j},k}\psi_{\mathrm{HU,j},k} \\ &+AF\left[\begin{array}{l}\sum_{i=1}^{N_{hot}}\sum_{j=1}^{N_{cold}}\sum_{k=1}^{N_{Hx}} a_{ijk} + b_{ijk}\left(A_{ijk}\right)^{\mathrm{C_k}}\psi_{ijk} + \sum_{i=1}^{N_{hot}}\sum_{k=1}^{N_{Hx}} a_{i,\mathrm{CU},k} + b_{i,CU,k}\left(A_{i,CU,k}\right)^{\mathrm{C_k}}\psi_{i,CU,k} \\ +\sum_{j=1}^{N_{cold}}\sum_{k=1}^{N_{Hx}} a_{HU,j,k} + b_{HU,j,k}\left(A_{HU,j,k}\right)^{\mathrm{C_k}}\psi_{\mathrm{HU,j},k}\end{array}\right]\end{aligned} \tag{1}$$

$$Q_{CU} = \sum_{i=1}^{N_{hot}}\sum_{k=1}^{N_{HX}} U_{i,CU,k} A_k LMTD_{i,CU,k}\psi_{i,CU,k} \tag{2}$$

$$Q_{HU} = \sum_{j=1}^{N_{cold}}\sum_{k=1}^{N_{HX}} U_{\mathrm{HU,j},k} A_k LMTD_{\mathrm{HU,j},k}\psi_{\mathrm{HU,j},k} \tag{3}$$

The Energy balance:

$$(\mathrm{T}_i^S - \mathrm{T}_i^T)\,\mathrm{MCp_i} = Q_{Process} + \sum_{i=1}^{N_{hot}}\sum_{k=1}^{N_{HX}} U_{i,CU,k} A_k LMTD_{i,CU,k}\psi_{i,CU,k} \qquad i \in H \tag{4}$$

$$(\mathrm{T}_j^S - \mathrm{T}_j^T)\,\mathrm{MCp}_j = Q_{Process} + \sum_{j=1}^{N_{cold}}\sum_{k=1}^{N_{HX}} U_{\mathrm{HU,j},k} A_k LMTD_{\mathrm{HU,j},k}\psi_{\mathrm{HU,j},k} \qquad j \in C \tag{5}$$

The constraints:

In this step, the following constraints can be specified:

- a limited number of new heat exchanger
- a limited number of re-sequencing
- a limited number of re-piping
- a limited m number of splitter

$$Q_{Process} = \sum_{i=1}^{N_{hot}}\sum_{j=1}^{N_{cold}}\sum_{k=1}^{N_{Hx}} U_{ijk}\ A_k\ \Delta T_{\underset{ijk}{LMTD}}\,\psi_{ijk} \tag{6}$$

$$\sum_{i=1}^{N_{hot}}\sum_{j=1}^{N_{cold}} \psi_{ijk} = 1 \tag{7}$$

$$\psi_{ijk} = \begin{cases} 1 & i \cup j \ in\ k_{Hx} \\ 0 & otherwise \end{cases}$$

$$\Delta T_{\substack{LMTD \\ ijk}} = \frac{\Delta T_1 - \Delta T_2}{Ln\left(\frac{\Delta T_1}{\Delta T_2}\right)} \quad ; \; m_i\ Cp_i\ \Delta T_i = m_j\ Cp_j\ \Delta T_j$$

$$\sum_{\substack{x=0 \\ x\neq k}}^{N_{Hx}} P_{xk} = 1 \quad ; \quad \sum_{\substack{m=0 \\ m\neq k}}^{N_{Hx}} L_{mk} = 1$$

$$P_{xy} = \begin{cases} 0, & No\ piping\ from\ \ x\ \rightarrow k \\ 1, & Have\ piping\ from\ x\ \rightarrow k \end{cases} ; \; L_{mk} = \begin{cases} 0, & No\ piping\ from\ m\ \rightarrow k \\ 1, & Have\ piping\ from\ m\ \rightarrow k \end{cases}$$

$Where$ $\quad P \equiv$ Piping for hot stream; $\; i \equiv$ hot stream

$L \equiv$ Piping for cold stream; $\; j \equiv$ cold stream

$N_{HX} \equiv$ Number of heat exchangers; k_{Hx} $\equiv$ heat exchanger

If $P_{xk} = 1$; $L_{mk} = 1$; $\quad$ x $\equiv m \equiv k \equiv heat\ exchanger\ index$

The Logarithmic mean temperature difference[ΔT_{LMTD}] in different match (ijk; HU,j ; i,CU) of superstructure for shell and tube exchanger for counter-current and co-current flow is given by:

$$\Delta T_{\substack{LMTD \\ i,j,k}} = \sum_{\substack{x=0 \\ x\neq k}}^{N_{Hx}} \sum_{\substack{m=0 \\ m\neq k}}^{N_{Hx}} \frac{\left(T_{\mathrm{i},x} - T_{j,k}\right) - \left(T_{\mathrm{i},k} - T_{j,m}\right)}{Ln\left(\frac{\left(T_{\mathrm{i},x} - T_{j,k}\right)}{\left(T_{\mathrm{i},k} - T_{j,m}\right)}\right)} P_{\mathrm{x},k}\, L_{mk} \qquad , i \in H\ , \; j \in C\ , \; k \in K \tag{8}$$

$$\Delta T_{\substack{LMTD \\ i,CU,K}} = \sum_{\substack{x=0 \\ x\neq k}}^{N_{Hx}} \sum_{\substack{m=0 \\ m\neq k}}^{N_{Hx}} \frac{\left(T_{i,x} - T_{\mathrm{CU},k}\right) - \left(T_{i,k} - T_{\mathrm{CU},m}\right)}{Ln\left(\frac{\left(T_{i,x} - T_{\mathrm{CU},k}\right)}{\left(T_{i,k} - T_{\mathrm{CU},m}\right)}\right)} P_{\mathrm{x},k}\, L_{mk} \qquad , \; i \in H \tag{9}$$

$$\Delta T_{\substack{LMTD \\ HU,j,k}} = \sum_{\substack{x=0 \\ x\neq k}}^{N_{Hx}} \sum_{\substack{m=0 \\ m\neq k}}^{N_{Hx}} \frac{\left(T_{\mathrm{HU},x} - T_{j,k}\right) - \left(T_{\mathrm{HU},k} - T_{j,m}\right)}{Ln\left(\frac{\left(T_{\mathrm{HU},x} - T_{j,k}\right)}{\left(T_{\mathrm{HU},k} - T_{j,m}\right)}\right)} P_{\mathrm{x},k}\, L_{mk} \qquad , \; j \in C \tag{10}$$

The area of heat exchanger in different match (ijk; HU, j; i, CU) of superstructure is given by:

$$A_{i,j,k} = \frac{q_{ijk}}{\Delta T_{LMTD_{i,j,k}} U_{i,j}} \qquad i \in H\ ,\ j \in C\ ,\ k \in K \tag{11}$$

$$A_{i,CU} = \frac{q_{i,CU}}{\Delta T_{LMTD_{i,CU}} U_{i,CU}} \qquad i \in H \tag{12}$$

$$A_{HU,j} = \frac{q_{HU,j}}{\Delta T_{LMTD_{HU,j}} U_{HU,j}} \qquad j \in C \tag{13}$$

When selected the overall heat transfer coefficient U will be calculated from the value specified in exchanger match is given by:

$$U_{i,j} = \left[\frac{1}{h_i} + \frac{1}{h_j}\right]^{-1} \tag{14}$$

$$U_{i,CU} = \left[\frac{1}{h_i} + \frac{1}{h_{CU}}\right]^{-1} \tag{15}$$

$$U_{HU,j} = \left[\frac{1}{h_{HU}} + \frac{1}{hj}\right]^{-1} \tag{16}$$

3.3 Process Simulation

The main objective of this study is to optimize the existing petrochemical units and operations. Design and or operating data from petrochemical plants was acquired and used to develop steady state models of all the units that were investigated. This was done by using the i-Heat process simulator software. The model was validated by estimating the deviation in the simulated data and operating data. If necessary, corrections and modifications were made to the simulation model to ensure a perfect representation of the operating process. The simulation model served as a computer/online version of the plant to study the influences of various process parameters on the plant and to carry out real-time optimization without having to revisit the plant for fresh data acquisition.

3.3.1 Overview of i-Heat

i-Heat is a software package used for the design and analysis of heat exchanger networks. The focal features contained within the i- heat are as follows:

- **Simulation**: for a given optimum network structure i-Heat will define the network heat exchanger performance and temperatures.
- **Optimization**: the program can automatically correct the branch flow and heat load distribution in the heat exchanger network to reach the specified design objective for revamp using optimization techniques.
- **Pinch analysis**: this analysis is used to detect the optimum network configuration for a given set of streams and provides the best design for energy and capital optimization.
- **Retrofit analysis**: This analysis is used to detect the network bottlenecks which limit energy recovery. Also, it can propose the best network structure modifications to overcome the limitations.
- **Automated design**: The program can automatically design a new network, according to the network temperatures and other specified constraints.
- **Select heat exchanger equipment automatically**: The program can help with the rating and recommend a list of suitable equipment for newly added exchangers and exchangers requiring additional space automatically.
- **Detailed simulation:** detailed calculation of exchangers can be performed given the geometry data.

The block diagram showing the application of i-Heat is shown in Fig.3.1.

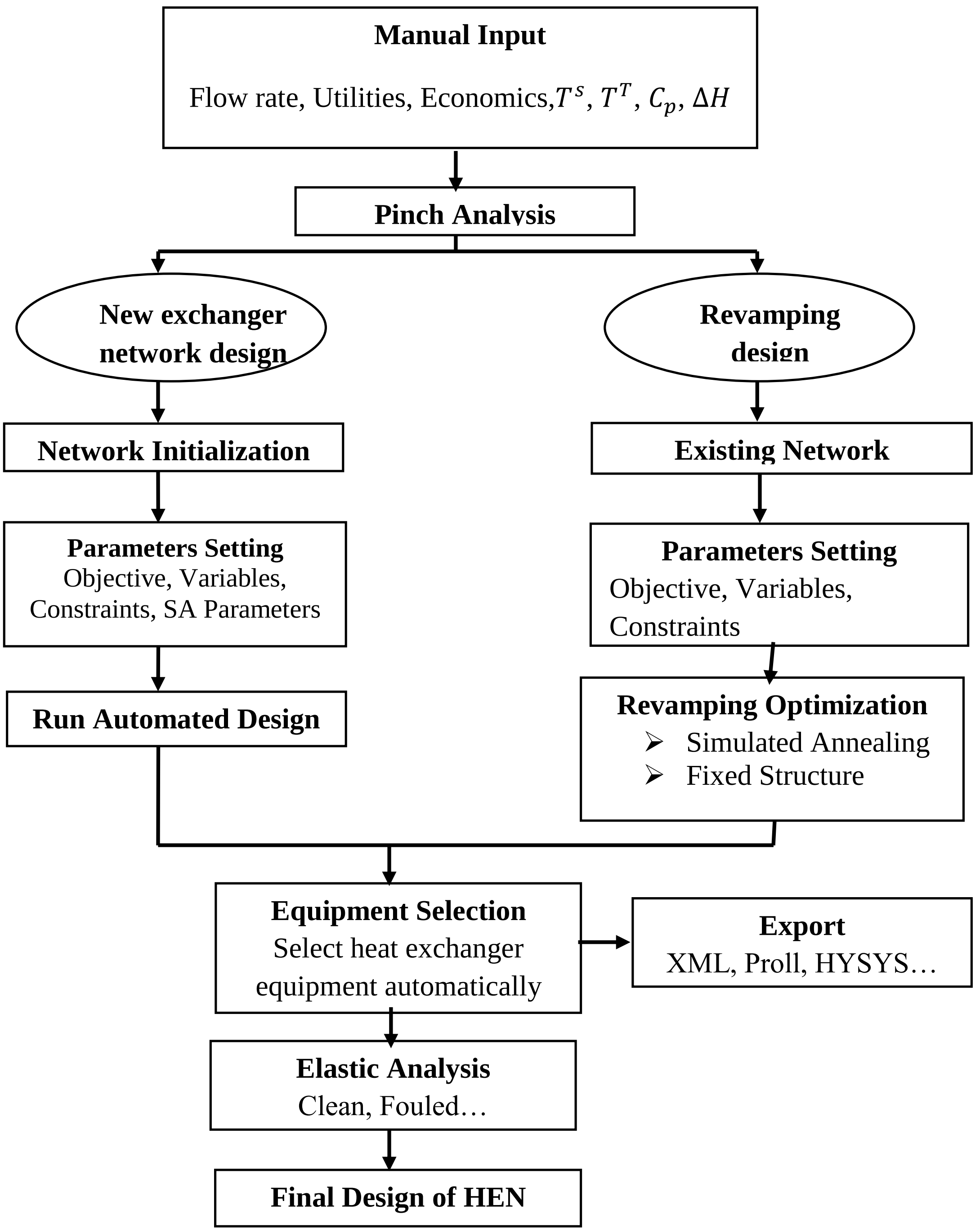

Figure 3.1: block diagram showing the application of i-Heat.

3.4 Network Retrofitting

Revamping a network involves adding additional area to existing matches and changing the network configuration, such as adding a new match and re-allocating existing matches. The problem is complex, with a lot of combinations of potential configuration modifications and trade-offs between energy and capital cost. Besides, there are various constraints that need to be considered, such as the maximum added area, space limits, added area threshold, etc. Therefore, a state-of-the-art revamping methodology is of great importance for generating a cost-effective and practical revamping solution.

The program provides an optimization-based revamping tool, which identifies the current network configuration bottleneck and the most appropriate places to install additional area that also generates the best energy recovery performance. When revamping a network, the tool should consider practical constraints such as maximum added area, exchangers that prohibit additional area, exchanger added area threshold, soft stream constraints, etc. Various revamping objectives can be considered, including the maximum temperature node, minimum utility cost, and total annualized cost. Besides, analyzing the energy effect of the background process change can be achieved by including the stream enthalpy change and heat capacity flow rate in the optimization.

The search for optimum conditions is based on two methods:

1. Fixed structured
2. Simulated Annealing.

3.4 1 Design Options:

1- **Simulated Annealing [SA]**

Stochastic optimization in which the structure being optimized is randomly moved from one state to another state by series of defined moves. The moves available depend on e nature of the structure being optimized. Many steps are followed, including:

a. **General**: In this step, an objective function is selected by minimizing Min Annualized Total Cost as follows:

Min Annualized Total Cost (TAC) = Annualized Energy cost + Annualized Capital Cost

The heat exchanger (HE) and utility cost are given by the following equations:

HE cost = $A_1 + B_1$ (area) ^{C_1}

Energy cost (Utility) = $A_2 + B_2$ (Duty) ^{C_2}

Where A represents fixed cost. B_1 is the heat exchanger cost per unit, dependent on the type of the material.

b. **Enthalpy change (ΔH):** the allowed process stream variations in revamping optimization. The function can be used to analyze the energy effect of background process change. Stream ΔH on the panel is selected. Then, this stream is included in optimization and the lower and upper bounds of enthalpy change ΔH are specified.

c. **Constraints and global network constraints**: In this step, the following constraints can be specified:
   - a limited number of new heat exchanger
   - a limited number of re-sequencing
   - a limited number of re-piping
   - a limited number of splitter

3.4.2 Configuration Analysis

Configuration analysis helps users to identify the bottlenecks in the current network configuration. Exchangers that cross pinch or criss-cross waste the heat transfer driving force and prohibit the energy recovery employing the same area. These structural bottlenecks can be overcome by modifying the network structure. These structural modifications can be classified as follows:

- **Re-sequencing:** this is the structural change when the location of an existing exchanger on a stream is altered such that the stream passes through the exchangers in an altered order. The exchanger is constrained to the same hot and cold streams. This is commonly the most useful modification such as safety, pressure rating, construction materials, etc., the network performance might be improved when the criss-cross exchangers are re-sequenced.

- **Re-piping**: this is the change where one side of an existing heat exchanger is connected to an altered stream.
- **New Exchanger:** this is the change when a new exchanger is added.
- **Stream Splitting:** this is the change where a stream is divided at a specific location to array the heat exchangers in parallel.

CHAPTER 4

Revamping Optimization and Thermodynamic Analysis of the Heat Exchanger Network in Propane Dehydrogenation Plant

4.1 Introduction

Considerable research efforts have been done over last decades and all over the world in the area of heat integration of chemical processes. With the occurrence of the energy crises in the 1970s, the wide attention of engineers, both in academia and in the industry, has been drawn to the economic utilization of energy in the chemical process industry. This situation, coupled with growing concern about the environment preservation, has resulted in large research efforts in energy optimization and process integration. Petrochemical industries are a major part of the economy in both the developed and developing countries. In these industries, attempts are continually made in order to maximize heat recovery, save energy, and minimize the impact of their activity on the environment[25]. Pinch analysis has been applied widely to study chemical processes, including those related to the petrochemical industry[3], [26]–[28]. Some research attempts considered the pinch analysis with constant parameters. On the other hand, some other attempts included flexible parameters to account for complex network configurations. These modifications still required structural adaptations with more practical approaches[5], [8], [29]–[31]. Al-Mutairi [19] has used the pinch analysis technique to modify the existing heat exchanger network in fluid catalytic cracking unit, by using the optimum ΔT_{min}. The proposed design has not only increased the heat recovery of the unit but also reduced the energy cost. Nowadays, there is an increasing attention toward the application of pinch concepts in the formulation of mathematical models in order to reduce the total annual operating cost [1], [20]. Floudas and Grossmann [10], [32] addressed this problem using various models and solvers. Indeed, Linear Programming (LP) was used to determine the lowest utility cost and location of the pinch points for each of the chosen parameters. In addition, Mixed Integer Linear Programming (MILP) was used for the determination of the minimum number of units. However, it should be noted that decomposing the Heat Exchanger Network (HEN) problem into separate targeting procedure (i.e., utility, energy, and area) does not guarantee that the total costs are minimized [15], [33], [34]. Mixed Integer Nonlinear Programming (MINLP) formulation [7], [9], [12] is efficient enough to capture the common features of the HEN and guarantee minimization of the total/overall cost of the network.

Both the global energy supply and the environmental situation require an improved management in the utilization of energy resources. Indeed, an improved petrochemical plant efficiency (i.e., optimum production at minimum energy consumption) is also capable of reducing carbon dioxide emissions associated with fossil-fuel combustion [35]. However, more research is still needed to for selecting the appropriate methodology for heat exchanger network (HEN) retrofitting by considering current networks, their constraints, and limitations [4].

In this paper, an improved heat exchanger networks (HENs) for optimum performance of the Propane Dehydrogenation (PDH) process at a minimum Total Annual Cost (TAC) is developed [36], [37]. Two techniques were adopted to optimize and retrofit the HEN, (1) the Simulated Annealing (SA) and (2) fixed structure technique using MINLP formulation. In the simulated annealing technique, the HEN is flexible and new heat exchanger can be added or removed. While in the fixed structure technique, the system allows a very limited space for a change of the number of heat exchangers in the network. The results of optimization can be applied to the PDH process by selecting the feasible revamp scheme. The results of our improved HENs will be compared to those of existing HEN.

Effective energy savings of the entire plant (i.e., processed-bottlenecking, investment cost reduction, process modification and total site planning) were presented as well in this study. In addition, information and relevant data on the thermodynamic indices for evaluating the performance of the investigated petrochemical units were provided including schemes and methods to reduce energy consumption and CO_2 emissions.

Here, the design and operating data, which were acquired from an existing process, were used to develop the steady state simulation of the process using i-heat process simulation software. Heat and mass balances data generated during the above phase were then used in order to optimize the PDH plant.

4.2 Methodology

The methodology was adopted in this study involves Revamping Optimization, simulation, and application of development models to solve optimization problems in the Petrochemical operations. Two techniques were implemented to achieve the desired goals.

First, the Simulated Annealing technique was applied to flexible heat exchanger network in order to minimize the Total Annual Cost (TAC).

The (TAC) is the objective function contains hot utility cost, cold utility cost, exchanger fixed duty and area cost is defined as follow, which is the modified version of the equation used by Al-Mutairi and Odejobi[1].

$$Min\ \mathrm{TAC} = C_{CU}\sum_{i=1}^{N_{hot}}\sum_{k=1}^{N_{HX}} U_{i,CU,k} A_k LMTD_{i,CU,k}\psi_{i,CU,k} + C_{HU}\sum_{j=1}^{N_{cold}}\sum_{k=1}^{N_{HX}} U_{\mathrm{HU,j},k} A_k LMTD_{\mathrm{HU,j},k}\psi_{\mathrm{HU,j},k}$$
$$+AF\left[\sum_{i=1}^{N_{hot}}\sum_{j=1}^{N_{cold}}\sum_{k=1}^{N_{Hx}} a_{ijk} + b_{ijk}\left(A_{ijk}\right)^{\mathrm{C_k}}\psi_{ijk} + \sum_{i=1}^{N_{hot}}\sum_{k=1}^{N_{Hx}} a_{i,\mathrm{CU},k} + b_{i,CU,k}\left(A_{i,CU,k}\right)^{\mathrm{C_k}}\psi_{i,CU,k} + \sum_{j=1}^{N_{cold}}\sum_{k=1}^{N_{Hx}} a_{HU,j,k} + b_{HU,j,k}\left(A_{HU,j,k}\right)^{\mathrm{C_k}}\psi_{\mathrm{HU,j},k}\right]$$
(1)

$$Q_{CU} = \sum_{i=1}^{N_{hot}}\sum_{k=1}^{N_{HX}} U_{i,CU,k} A_k LMTD_{i,CU,k}\psi_{i,CU,k} \qquad (2)$$

$$Q_{HU} = \sum_{j=1}^{N_{cold}}\sum_{k=1}^{N_{HX}} U_{\mathrm{HU,j},k} A_k LMTD_{\mathrm{HU,j},k}\psi_{\mathrm{HU,j},k} \qquad (3)$$

The Energy balance:

$$(\mathrm{T}_i^S - \mathrm{T}_i^T)\,\mathrm{MCp_i} = Q_{Process} + \sum_{i=1}^{N_{hot}}\sum_{k=1}^{N_{HX}} U_{i,CU,k} A_k LMTD_{i,CU,k}\psi_{i,CU,k} \qquad i \in H \qquad (4)$$

$$(\mathrm{T}_j^S - \mathrm{T}_j^T)\,\mathrm{MCp}_j = Q_{Process} + \sum_{j=1}^{N_{cold}}\sum_{k=1}^{N_{HX}} U_{\mathrm{HU,j},k} A_k LMTD_{\mathrm{HU,j},k}\psi_{\mathrm{HU,j},k} \qquad j \in C \qquad (5)$$

The constraints:

In this step, the following constraints can be specified:

- a limited number of new heat exchanger
- a limited number of re-sequencing
- a limited number of re-piping
- a limited m number of splitter

$$Q_{Process} = \sum_{i=1}^{N_{hot}} \sum_{j=1}^{N_{cold}} \sum_{k=1}^{N_{Hx}} U_{ijk}\, A_k\, \Delta T_{\underset{ijk}{LMTD}}\, \psi_{ijk} \qquad (6)$$

$$\sum_{i=1}^{N_{hot}} \sum_{j=1}^{N_{cold}} \psi_{ijk} = 1 \qquad (7)$$

$$\psi_{ijk} = \begin{cases} 1 & i \cup j \; in \; k_{Hx} \\ 0 & otherwise \end{cases}$$

$$\Delta T_{\underset{ijk}{LMTD}} = \frac{\Delta T_1 - \Delta T_2}{Ln\left(\dfrac{\Delta T_1}{\Delta T_2}\right)} \quad ; \; m_i \; Cp_i \; \Delta T_i = m_j \; Cp_j \; \Delta T_j$$

$$\sum_{\substack{x=0 \\ x\neq k}}^{N_{Hx}} P_{xk} = 1 \quad ; \quad \sum_{\substack{m=0 \\ m\neq k}}^{N_{Hx}} L_{mk} = 1$$

$$P_{xy} = \begin{cases} 0, & No\ piping\ from\ \ x \ \to k \\ 1, & Have\ piping\ from\ x \ \to k \end{cases} \; ; L_{mk} = \begin{cases} 0, & No\ piping\ from\ m \ \to k \\ 1, & Have\ piping\ from\ m \ \to k \end{cases}$$

$Where$ $P \equiv$ Piping for hot stream; $i \equiv$ hot stream

$L \equiv$ Piping for cold stream; $j \equiv$ cold stream

$N_{HX} \equiv$ Number of heat exchangers; k_{Hx} $\equiv$ heat exchanger

If $P_{xk} = 1$; $L_{mk} = 1$; x $\equiv m \equiv k \equiv heat\ exchanger\ index$

The Logarithmic mean temperature difference[ΔT_{LMTD}] in different match (ijk; HU,j ; i,CU) of superstructure for shell and tube exchanger for counter-current and co-current flow is given by:

$$\Delta T_{\substack{LMTD\\ i,j,k}} = \sum_{\substack{x=0\\ x\neq k}}^{N_{Hx}} \sum_{\substack{m=0\\ m\neq k}}^{N_{Hx}} \frac{\left(T_{\mathrm{i},x} - T_{j,k}\right) - \left(T_{\mathrm{i},k} - T_{j,m}\right)}{Ln\left(\dfrac{\left(T_{\mathrm{i},x} - T_{j,k}\right)}{\left(T_{\mathrm{i},k} - T_{j,m}\right)}\right)} P_{\mathrm{x},k}\, L_{mk} \qquad , i \in H\ ,\ j \in C\ ,\ k \in K \tag{8}$$

$$\Delta T_{\substack{LMTD\\ i,CU,K}} = \sum_{\substack{x=0\\ x\neq k}}^{N_{Hx}} \sum_{\substack{m=0\\ m\neq k}}^{N_{Hx}} \frac{\left(T_{i,x} - T_{\mathrm{CU},k}\right) - \left(T_{i,k} - T_{\mathrm{CU},m}\right)}{Ln\left(\dfrac{\left(T_{i,x} - T_{\mathrm{CU},k}\right)}{\left(T_{i,k} - T_{\mathrm{CU},m}\right)}\right)} P_{\mathrm{x},k}\, L_{mk} \qquad ,\ i \in H \tag{9}$$

$$\Delta T_{\substack{LMTD\\ HU,j,k}} = \sum_{\substack{x=0\\ x\neq k}}^{N_{Hx}} \sum_{\substack{m=0\\ m\neq k}}^{N_{Hx}} \frac{\left(T_{\mathrm{HU},x} - T_{j,k}\right) - \left(T_{\mathrm{HU},k} - T_{j,m}\right)}{Ln\left(\dfrac{\left(T_{\mathrm{HU},x} - T_{j,k}\right)}{\left(T_{\mathrm{HU},k} - T_{j,m}\right)}\right)} P_{\mathrm{x},k}\, L_{mk} \qquad , j \in C \tag{10}$$

The area of the heat exchanger in a different match (ijk; HU, j; i, CU) of the superstructure is given by:

$$A_{\mathrm{i,j,k}} = \frac{q_{ijk}}{\Delta T_{\substack{LMTD\\ i,j,k}} U_{i,\mathrm{j}}} \qquad i \in H\ ,\ j \in C\ ,\ k \in K \tag{11}$$

$$A_{\mathrm{i,CU}} = \frac{q_{i,CU}}{\Delta T_{\substack{LMTD\\ i,\mathrm{CU}}} U_{i,\mathrm{CU}}} \qquad i \in H \tag{12}$$

$$A_{\mathrm{HU,j}} = \frac{q_{HU,j}}{\Delta T_{\substack{LMTD\\ \mathrm{HU},j}} U_{\mathrm{HU,j}}} \qquad j \in C \tag{13}$$

When selected the overall heat transfer coefficient U will be calculated from the value specified in exchanger match is given by:

$$\mathrm{U}_{\mathrm{i,j}} = \left[\frac{1}{h_i} + \frac{1}{h_j}\right]^{-1} \tag{14}$$

$$U_{i,CU} = \left[\frac{1}{h_i} + \frac{1}{h_{CU}}\right]^{-1} \quad (15)$$

$$U_{HU,j} = \left[\frac{1}{h_{HU}} + \frac{1}{hj}\right]^{-1} \quad (16)$$

Where U_K the overall heat transfer coefficient, A_k is area of heat exchanger, ΔT_{LMTD} is the Log Mean Temperature Difference (LMTD) for shell and tube exchanger for counter-current and co-current flow.

Second, the fixed structure method was applied to get the minimum Total Annual Cost based on

1. a minimum change in the structure of HEN,
2. different process stream rate, and
3. varying process stream heat exchanger duties and heat exchanger areas.

At a final stage of our study, we compare the two techniques and select the best based on the lowest Total Annual Cost criteria.

4.3 Process Description

4.3.1 Overall Description

The Propane Dehydrogenation Unit (PDH) produce a polymer grade propylene product from an LPG / propane feed stream. The Process Block Diagram of PDH unit is given in figure (4.1).

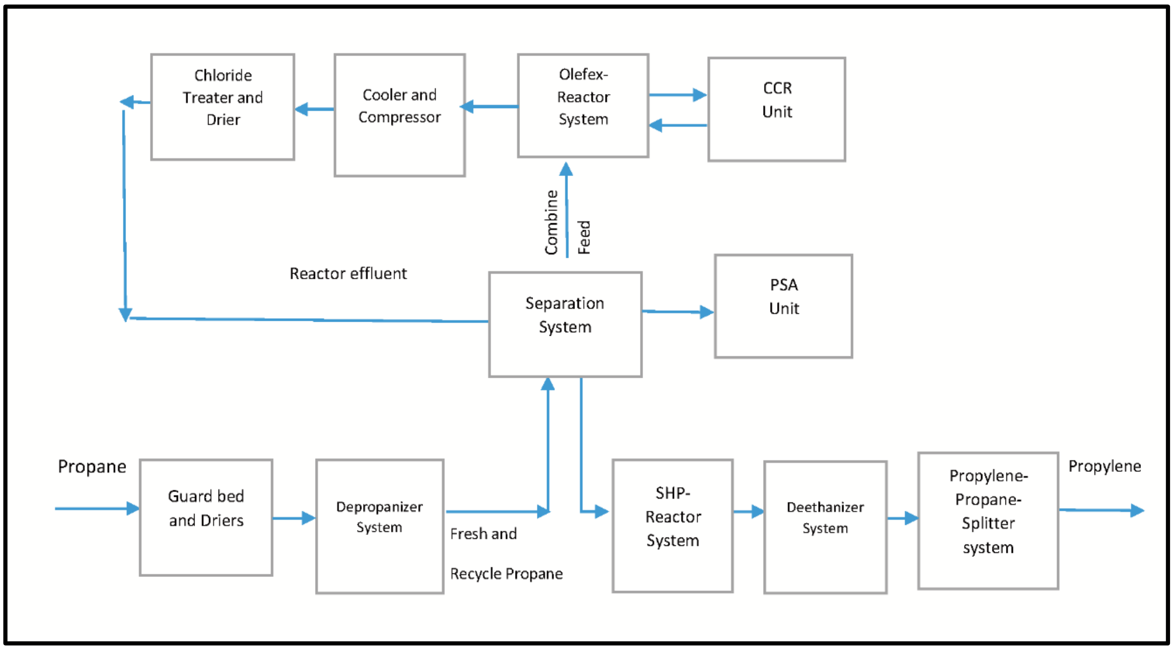

Figure 4.1: Process Block Diagram for the Propane Dehydrogenation Unit (PDH)

The unit's initial yield is 50 metric tons per hour (t/h). While only 47.5 t/h are produced at the end of run conditions. Propylene production is further increased with the extra processing of 1.5 t/h of recovered material from the downstream polypropylene production unit.

The propylene product from the PDH plant is later processed in the downstream process unit for the production of polypropylene. The entire PDH-unit consists of the process sections and the supporting systems (i.e., cooling and heating systems …).

4.3.2 Existing heat exchanger network

The current process consists of 25 heat exchangers. There is a total of 50 streams. 12 cold process streams and 13 hot process streams as well as 12 hot and 13 cold utilities streams. The hot exchanger's utility uses either HP, LP steam or hot water while the cold utilities heat exchangers use either cooling water or air supply. The details of all the streams and utilities are listed in Table (4.1) and (4.2), respectively. The grid diagram of existing heat exchanger network of (PDH) is shown in figure (4.2).

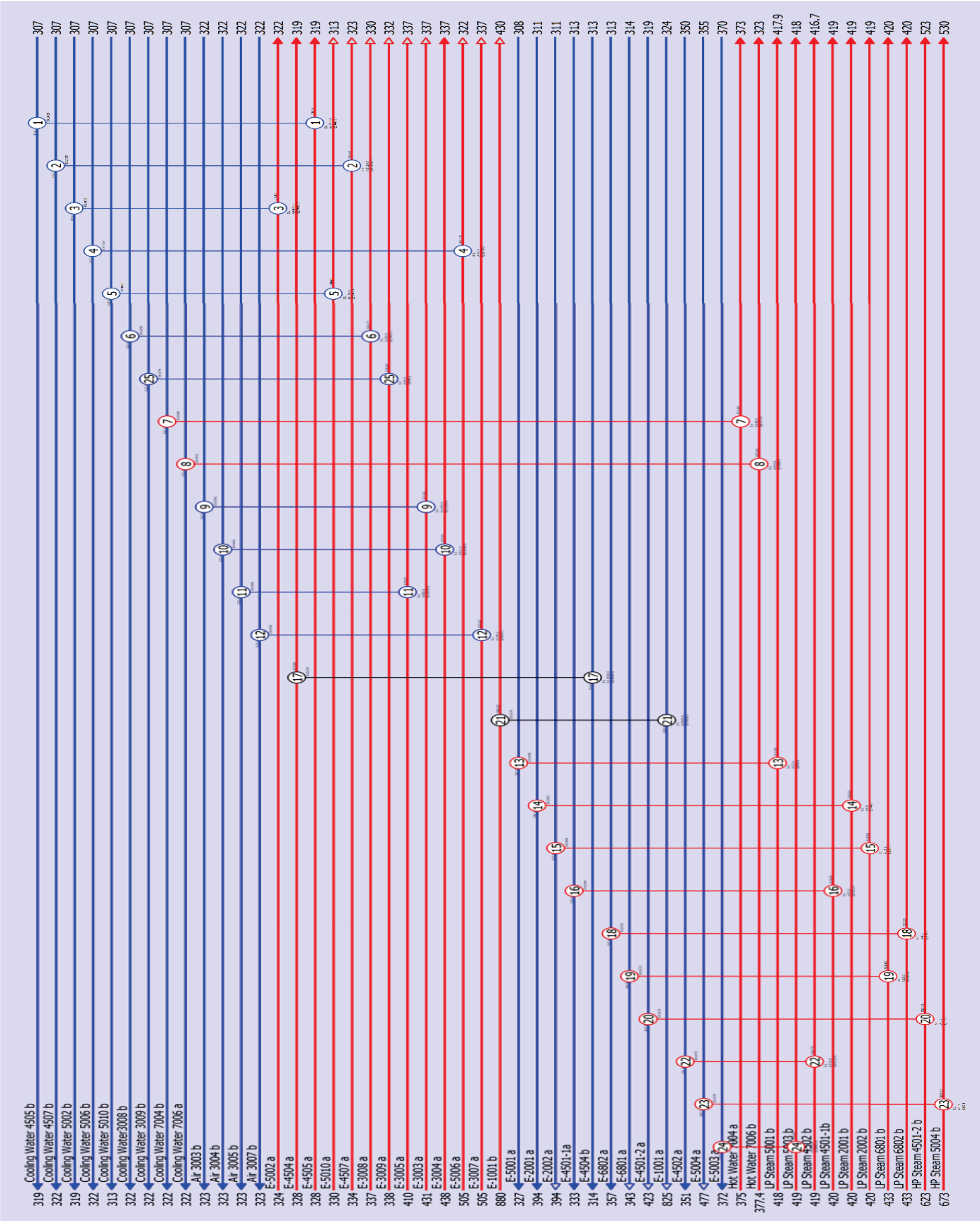

Figure 4.2: Grid diagram of existing heat exchanger network of propene dehydrogenation unit (PDH)

Table 4.1: The hot and cold stream for existing design

No.	Stream	Flow rate [kg/h]	T^S [K]	T^T [K]	MCp [kW/K]	DH [kW]	FR [(m²•K)/kW]	Inlet Pressure [bar]	Cp [kJ/(kg•K)]
1	E-4501-1a	166321	313	333	125.00	2500	0.00026	37	2.71
2	E-4504 a	660913	328	319	6262.22	56360	0.0001	18.5	34.11
3	E-4504 b	2207244	313	314	56360.00	56360	0.0001	13.2	91.92
4	E-4505 a	189761	328	319	1797.78	16180	0.00026	18.2	34.11
5	E-4507 a	303802	334	323	1618.18	17800	0.00026	29.3	19.18
6	E-5001 a	6414	308	327	31.58	600	0.00026	1.01325	17.72
7	E-5002 a	343793	324	322	13815.00	27630	0.00026	16.9	144.66
8	E-5003 a	1954991	370	372	14600.00	29200	0.00035	17.8	26.89
9	E-5004 a	201	355	477	0.41	50	0.00026	4	7.34
10	E-5006 a	6414	505	322	6.99	1280	0.00026	17.4	3.93
11	E-5010 a	64142	330	313	58.24	990	0.00026	24.1	3.27
12	E-1001 a	174999	324	825	149.16	74730	0.00035	3	3.07
13	E-1001 b	175623	880	430	166.07	74730	0.00035	0.38	3.40
14	E-3008 a	176442	337	330	110.00	770	0.00018	13.3	2.24
15	E-3009 a	11840	338	332	42.50	255	0.00018	3.49	12.92
16	E-2001 a	429	311	394	1.69	140	0.00018	4.1	14.15
17	E-2002 a	967	311	394	2.29	190	0.00018	4.3	8.52
18	E-3003 a	175912	431	337	121.60	11430	0.0007	0.22	2.49
19	E-3004 a	176442	438	337	124.26	12550	0.00027	4.17	2.54
20	E-3005 a	176442	410	337	123.84	9040	0.00036	13.5	2.53
21	E-3007 a	11840	505	337	14.46	2430	0.0007	3.67	4.40
22	E-4502 a	614811	350	351	24210.00	24210	0.00035	30.3	141.76

23	E-4501-2 a	250	319	423	1.06	110	0.00026	2.1	15.23
24	E-6801 a	25560	314	343	24.14	700	0.0002	6.8	3.40
25	E-6802 a	6000	313	357	6.17	271.3	0.0002	6	3.70

Table 4.2: The hot and cold utilities for existing design

	Utility	T^S [K]	T^T [K]	FR [(m²•K)/kW]	Flow rate [kg/h]	Cp [kJ/(kg•K)]	Inlet Pressure [bar]
1	LP Steam 4501-1b	420	419	9.00E-05	4229	2.32	3.2
2	Cooling Water 4505 b	307	319	0.00018	1742562	4.18	4.5
3	Cooling Water 4507 b	307	322	0.00018	1022789	4.18	4.5
4	LP Steam 5001 b	418	417.9	9.00E-05	1015	2.32	2.32
5	Cooling Water 5002 b	307	319	0.00018	1984240	4.18	4.5
6	LP Steam 5003 b	419	418	0.00018	49338	2.32	3.2
7	HP Steam 5004 b	673	530	9.00E-05	71	3.73	43.5
8	Cooling Water 5006 b	307	322	0.00018	73550	4.18	4.5
9	Cooling Water 5010 b	307	313	0.00018	64142	4.18	4.5
10	Cooling Water3008 b	307	322	0.00018	43962	4.18	4.5
11	Cooling Water 3009 b	307	322	0.00018	14673	4.18	4.51
12	LP Steam 2001 b	420	419	0.0001	242	2.32	3.2
13	LP Steam 2002 b	420	419	0.0001	317	2.32	3.2
14	Air 3003 b	322	323	0.00035	47162185	1	1.01
15	Air 3004 b	322	323	0.00035	45180000	1	1.01
16	Air 3005 b	322	323	0.00035	38086334	1	1.01
17	Air 3007 b	322	323	0.00035	9464843	1	1.01
18	LP Steam 4502 b	419	416.7	9.00E-05	40874	2.19	3.2
19	HP Steam 4501-2 b	623	523	9.00E-05	270	3.7	41
20	LP Steam 6801 b	433	420	0.0001	1180	2.32	3.5
21	LP Steam 6802 b	433	420	0.0001	460	2.32	3.5
22	Hot Water 7004 A	375	373	0.0001	9787	4.18	0.1
23	Cooling Water 7004 b	307	322	0.00018	354595	4.18	4

24	Hot Water 7006 a	377.4	323	0.0001	6300	4.18	0.2
25	Cooling Water 7006 b	307	322	0.00018	391630	4.18	4

4.4 Cost Data and Operation Cost

In order to take a decision about the economic feasibility of the revamp plant, an economical evaluation must be performed. The typical operation time is 8600 hours, while the plant is assumed to have a lifetime of 5 years. The annual interest rate is assigned to be roughly 6%. The mathematical formula, used to determine the annualization factor [38], is given by:

$$\text{Annualization factor} = R\,(1+R)^{n} / (1+R)^{n} - 1$$

The operation costs are mainly related to the consumption of fuel for heat generation. This heat is necessary for steam generation in order to supply the hot streams. Moreover, the cost associated with cold utilities is also included, but it is on the lower side in comparison with that required for the hot utilities. Table (4.3) below lists the cost data of the utilities from the local company in Saudi Arabia.

Table 4.3: The fuel price of the hot and cold utilities

Hot Utility	**Fuel price**
HP steam	7.2 US $/ton
LP steam	6.2 US $/ton
Hot water	2 US $/ton
Cold Utility	

Cooling water	0.2 US $/ton
Air cooling	0.2 $/KWh

Capital cost

The heat exchanger (HE) and utility cost are given by the following equations:

HE cost = $A_1 + B_1 (\text{area})^{C_1}$

Capital cost (Utility) = $A_2 + B_2 (\text{Duty})^{C_2}$

Total Annualized Cost (TAC) = Annualized Energy cost + Annualized Capital Cost

Where A represents the fixed cost.

Here, we take A_1 and $A_2 = 0$ since we consider only revamping of an existing design. A_1 and A_2 will be nonzero in the case where a new heat exchanger is added after retrofitting.

B_1, the heat exchanger cost per unit, depends on the type of the material as indicated in Table (4.4).

Table 4.4: The Price of heat exchanger's material

Material	Density kg/m3	Material price $/t	Thermal conductivity W/(m.k)
CS	7850	530	46.7
SS	7930	550	19

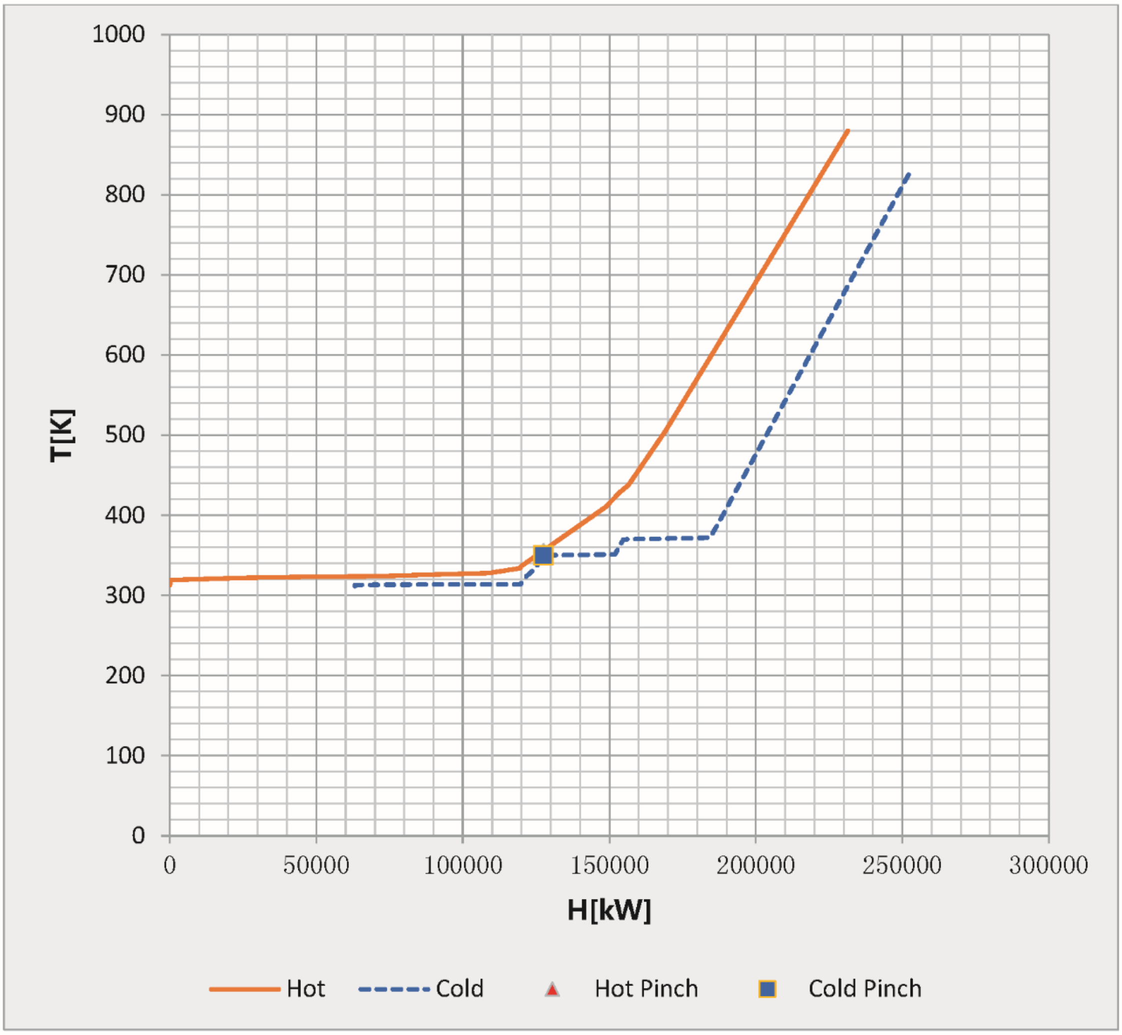

Figure 4.3: Composite Curves for existing Heat Exchanger Network of (PDH)

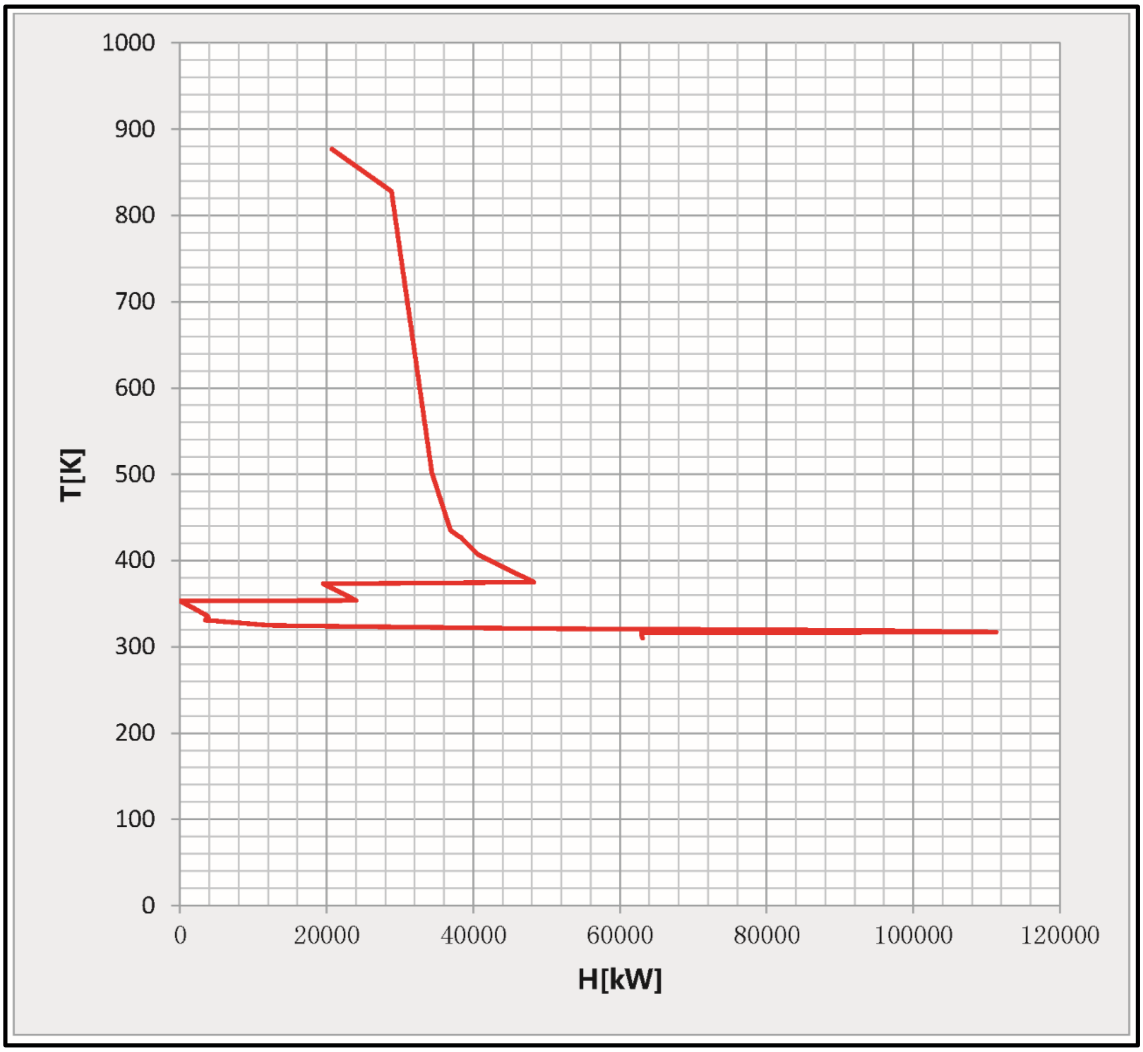

Figure 4.4: The Existing Heat Exchanger Network Grand Composite Curve of (PDH)

4.5 Result and Discussion

4.5.1 Pinch Analysis

Pinch Analysis was used to carry on the optimization of the process. It was found that the hot and cold utility demands of the existing network were 72.54 MW and 128.53 MW, respectively as shown in the composite curve in figure (4.3). In addition, the grand composite curve (GCC) is presented in figure (4.4) showing the variation of utilities selected.

4.5.2 Pinch configuration

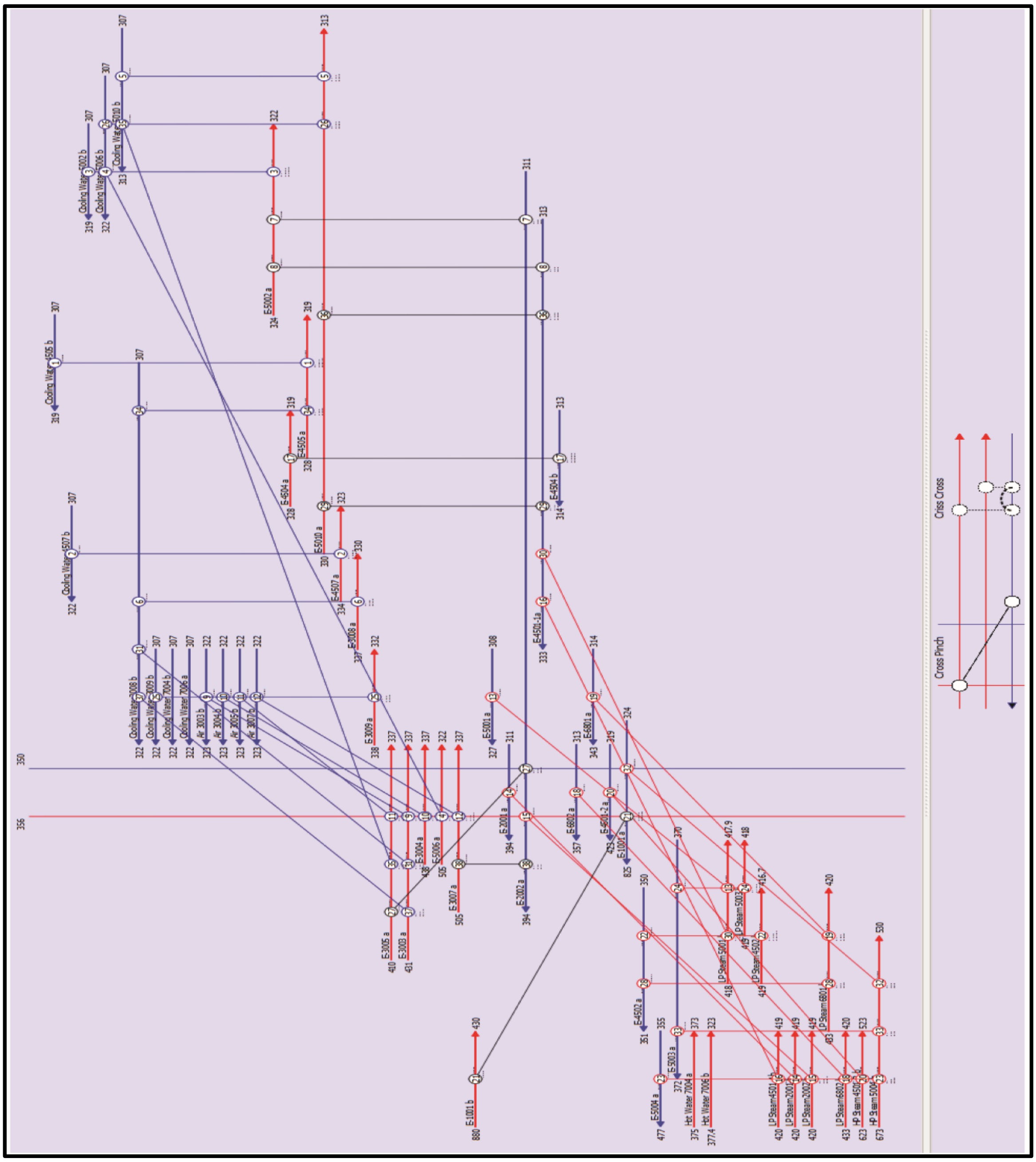

Figure 4.5: A1. Pinch configuration for existing network

Indeed, when pinch analysis was applied to the current heat exchangers network, 15 heat exchangers out of the 25 (EX1074, EX1073, EX1072, EX1071, EX1069, EX1068, EX1067,EX1066, EX1065, EX1064, EX1063, EX1062, EX1061, EX1060, EX1057) crossed the pinch temperature as shown in Table (4.5) and figure (4.A1 support document)

Table 4.5: Heat exchangers that across the pinch and cross pinch duty for existing network

No	Name	Duty [kW]	Cross pinch duty[kW]	Hot side (Pinch Temperature:356)			
				Stream Name	MCp [kW/K]	Inlet Temp [K]	Outlet Temp [K]
21	EX1074	79936.1	3878.2	E-1001 b	166.067	880	398.651
20	EX1073	112.014	32.7885	HP Steam 4501-2 b	1.12014	623	523
19	EX1072	1226.97	868.966	LP Steam 6801 b	94.3825	433	420
18	EX1071	276.205	228.139	LP Steam 6802 b	21.2465	433	420
16	EX1069	2551.95	2551.95	LP Steam 4501-1b	2551.95	420	419
15	EX1068	235.285	89.2771	LP Steam 2002 b	235.285	420	419
14	EX1067	140	65.7831	LP Steam 2001 b	140	420	419
13	EX1066	600	600	LP Steam 5001 b	6000	418	417.9
12	EX1065	2629.12	2155.18	E-3007 a	14.4643	505	323.233
11	EX1064	10579.5	6687.12	E-3005 a	123.836	410	324.568
10	EX1063	12550	10189.1	E-3004 a	124.257	438	337
9	EX1062	13100.6	9119.68	E-3003 a	121.596	431	323.261
8	EX1061	7068.44	2780.6	Hot Water 7006 b	129.935	377.4	323
7	EX1060	6171	6171	Hot Water 7004 a	3085.5	375	373
4	EX1057	1376.58	1042.19	E-5006 a	6.99454	505	308.192

4.6 Revamping Optimization

After identifying the bottleneck from the pinch configuration, we attempt to revamp the networks using the available optimization techniques. The optimum choice is obtained by adding new additional area, new matches, re-allocating existing matches within the various constrains such as maximum added area, space limits. We will search for a cost effective as well as a practical revamping solution.

The search of this optimum is based on two methods:

3. fixed structured and
4. Simulated Annealing.

Figure (4.5) shows the detailed block diagram of the retrofitting procedure using the i-heat commercial software.

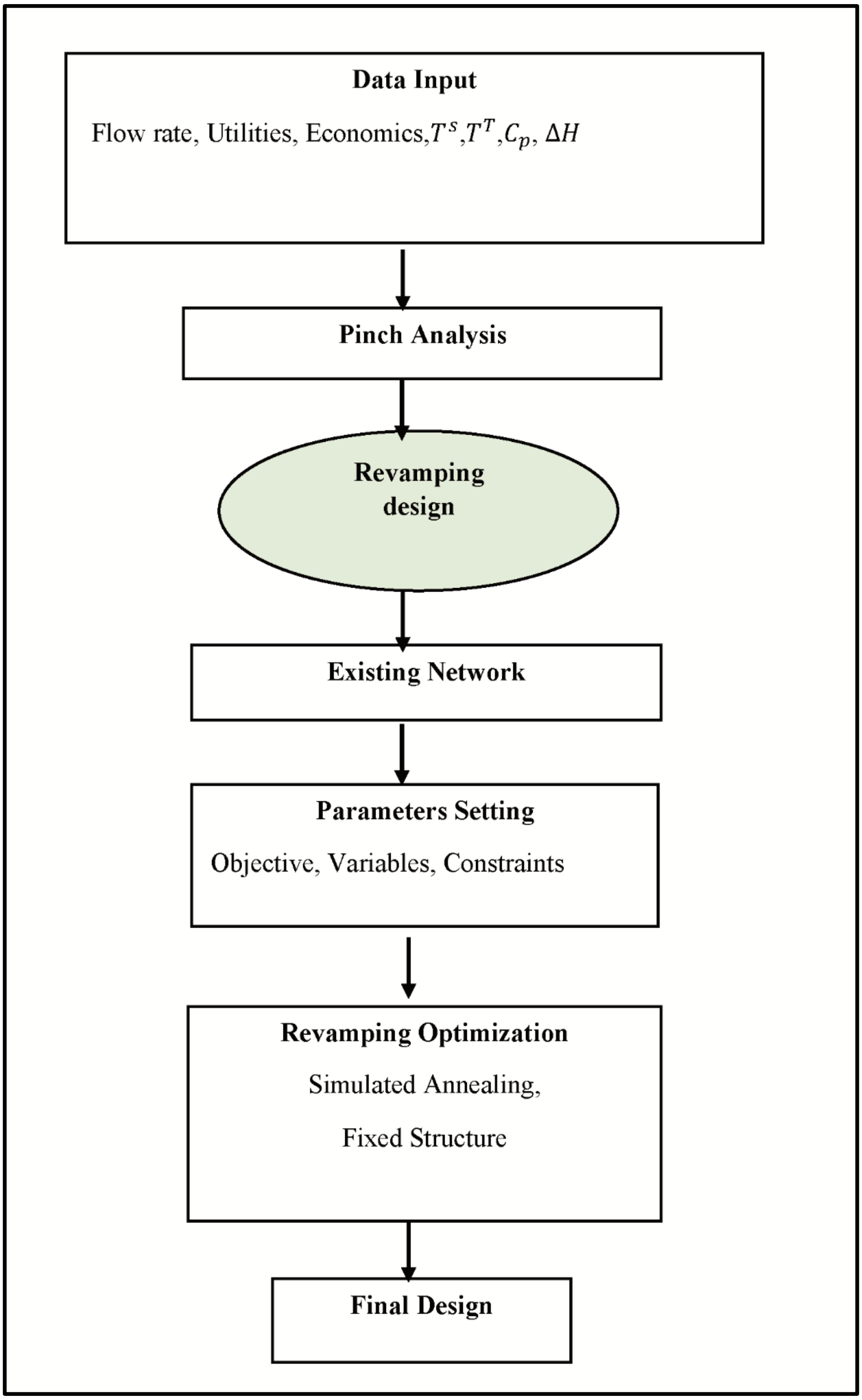

Figure 4.6: The block diagram showing Retrofitting Procedure

4.7 Design Options:

4.7.1 Simulated Annealing [SA]

Stochastic optimization in which the structure being optimized is randomly moved from one state to another state by series of defined moves. The moves available depend on the nature of structure being optimized.

The flow diagram showing the steps of the generalized method for HEN synthesis, using stochastic optimization is depicted in Figure (4.6).

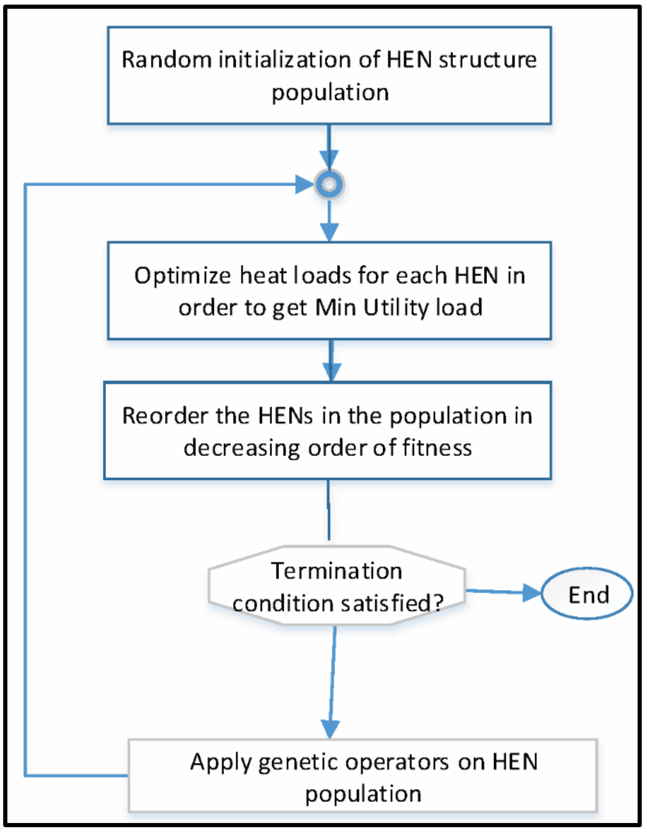

Figure 4.7: Flow diagram describing the general framework and minimum utility load optimal synthesis of HEN using Simulation Annealing.

In order to carry out the optimization of the HEN, we did some modifications according to the constraints by the pinch configuration. Thirteen heat exchangers were added namely 26, 27, 28, 29, 30, 31, 32, 33,34,35,36, 37, and 38. Thus, the total number of heat exchangers in the retrofit design become 38. Also, the areas of HE-10 (E-3004) and

HE-13 (5004) were increased. Finally, ten heat exchangers were re-sequenced namely [1-2-3-4-5-6-9-11-25] and two heat exchangers were re-piped namely [7-8]. The details of the re-piped streams are listed in Table (4.6). The comparison of the existing and retrofit HEN is shown schematically in Figure (4.7).

Table 4.6: The re-piped HEs

Name of re-piped HEs	Existing (before revamping) the connection between:	After revamping the connection between:
HE 7	Stream E-7004 b (cold water) and stream E-7004 a (hot water)	Stream E-5002 a and stream E-2002 a
HE 8	Stream E-7006 a (cold water) and stream E-7006 b (hot water)	Stream E-5002 a and stream 4501-1a

Table 4.7: Heat exchangers that across the pinch and cross pinch duty for Simulated Annealing

Name	Duty [kW]	Cross pinch duty[kW]	Hot side (Pinch Temperature:356)			
			Stream Name	F*Cp [kW/K]	Inlet Temp[K]	Outlet Temp[K]
EX12073	172.8	172.8	E-3003 a	121.6	431.0	429.6
EX21080	82.6	82.6	E-3005 a	123.8	409.6	409.0
EX40014	0.4	0.4	HP Steam 5004 b	0.4	531.1	530.0
EX16551	253.7	253.7	E-3003 a	121.6	429.6	427.5
EX14587	425.6	425.6	LP Steam 5001 b	10256.2	418.0	418.0
EX23740	45.8	45.8	E-3005 a	123.8	410.0	409.6
EX1074	74729.6	3877.8	E-1001 b	166.1	880.0	430.0
EX1073	110.0	32.8	HP Steam 4501-2 b	1.1	623.0	523.0
EX1072	700.0	700.0	LP Steam 6801 b	54.7	432.8	420.0
EX1071	271.3	228.1	LP Steam 6802 b	20.9	433.0	420.0
EX1069	1586.4	1586.4	LP Steam 4501-1b	1586.4	420.0	419.0
EX1068	142.5	43.5	LP Steam 2002 b	142.5	420.0	419.0
EX1067	140.0	65.8	LP Steam 2001 b	140.0	420.0	419.0
EX1066	600.0	600.0	LP Steam 5001 b	10256.2	418.0	417.9
EX1065	2428.3	2153.5	E-3007 a	14.5	504.9	337.0
EX1064	8911.7	6558.8	E-3005 a	123.8	409.0	337.0
EX1063	12550.0	10189.1	E-3004 a	124.3	438.0	337.0
EX1062	11003.5	8693.2	E-3003 a	121.6	427.5	337.0
EX1057	1280.0	1042.2	E-5006 a	7.0	505.0	322.0

The Simulated Annealing method was able to reduce the HEs duty from 46.4 MW to 36.6 MW due to abandoning the HEs that were crossing the pinch in the existing network. The HEs and their analysis are presented in figure (4.A2 support document) and Table (4.7)

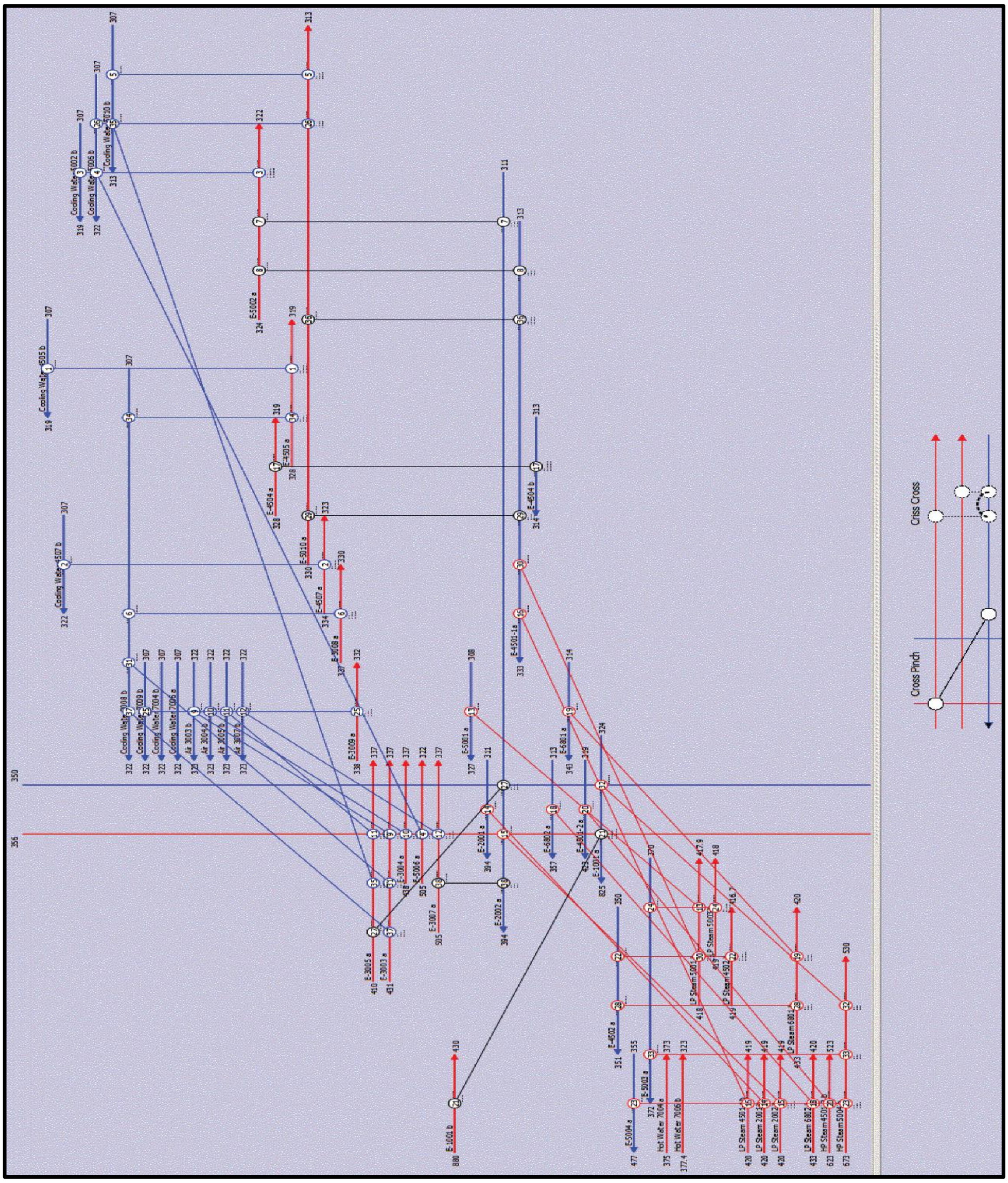

Figure 4.8: A2. Pinch configuration for Simulated Annealing (SA)

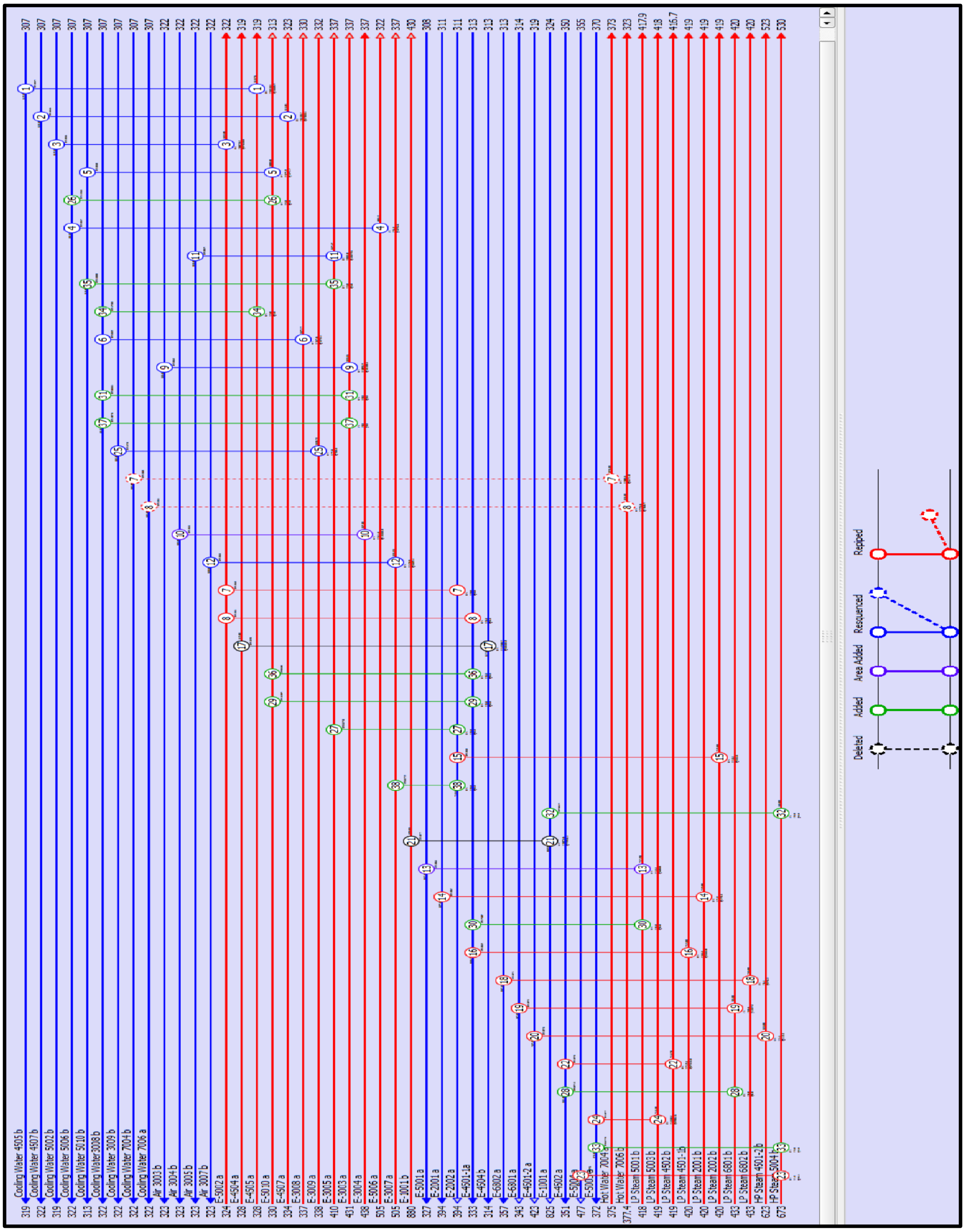

Figure 4.9: Superstructure showing a comparison between the existing network and the retrofitted one by Simulated Annealing (SA)

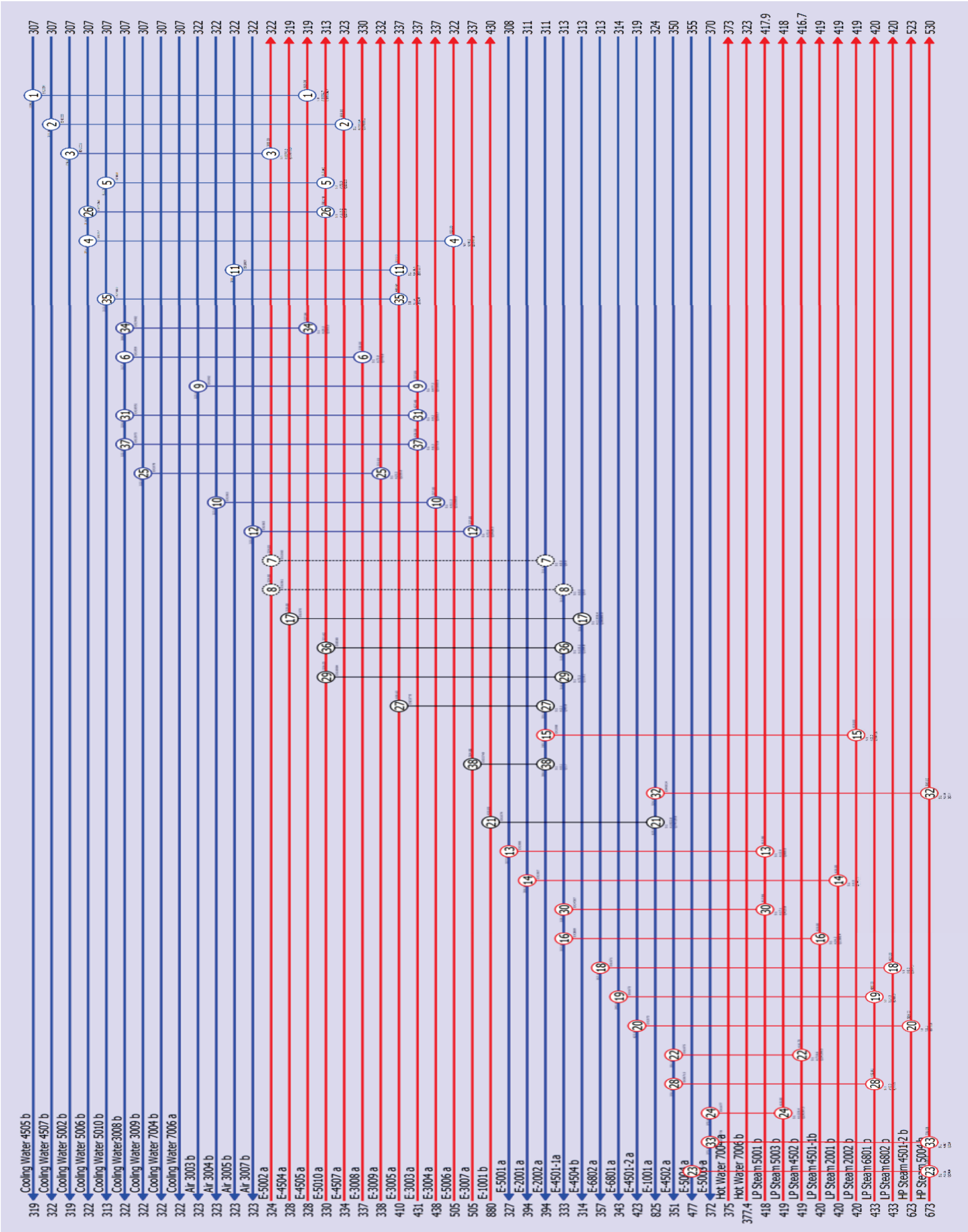

Figure 4.10: The final design after optimization using Simulated Annealing [SA]

Table 4.8: Summary of result of Simulated Annealing [SA] (comparison between the Existing Network and Retrofitting network)

Items		Retrofitting network	Existing Network
ΔTmin	K	6.00	6.00
Hot utility duty	MW	57.44	72.54
Cold utility duty	MW	99.82	128.53
Number of heat exchangers		38.00	25.00
Added area	m^2	29678.3	36236.3
Used area	m^2	29678.3	36236.3
Total Annualized cost	MM$/y	120.42	141.07
Total capital cost	MM$/y	7.22	9.18
Total operating cost	MM$/y	113.20	131.89
Hot utility operating cost	MM$/y	26.78	32.94
Cold utility operating cost	MM$/y	67.55	76.97
Another operating cost	MM$/y	18.87	21.98

The results of the retrofit design by simulated annealing are summarized in Table (4.8) and shown in figure (4.8). The Total Annual Cost is of the network was reduced from 141.07 MM$/y to 120.42 MM$/y, which constitutes a saving of up to 14.64%.

4.7.2 Fixed structure:

The second optimization technique that was employed, to minimize the annual operating cost, is called fixed structure method. In this method, minimum modification of the heat exchanger network has been considered by varying: (1) hot and cold stream rates, (2) heat exchangers duties and areas.

We found that 16 heat exchangers cross the pinch as shown in figured (A3 support document) [New HX1080, EX1074, EX1073, EX1072, EX1071, EX1069, EX1068, EX1067, EX1066, EX1065, EX1064, EX1063, EX1062 and EX1061] as shown in Table (4.9). Moreover, the criss-cross heat transfer has been found for the hot stream E-4505 heat exchanger as shown in Table (4.10).

Indeed, when Fixed Structured was applied, to the current heat exchangers network, the duty of HE, that crossed the pinch, was reduced.

The modified superstructure is shown in figure (4.9). This superstructure, obtained with the fixed structure method, reduces the annual operating cost. Two HEs are added to the system: (1) HX1079 matching between E-4504a and cooling water 4505 b, and (2) HX1080 matching between E-1001 b and cooling water E-4505b with areas of 17670.5 m2, 1055.2 m2, respectively. The final design for the fixed structure shown in figure (4.10) and table (4.11) summarizes the reduction of Total Annual Cost using this technique. In effect, the TAC is reduced from 141.89 MM$/y to 130.00 MM$/y (i.e., up to 7.8 % saving is achieved).

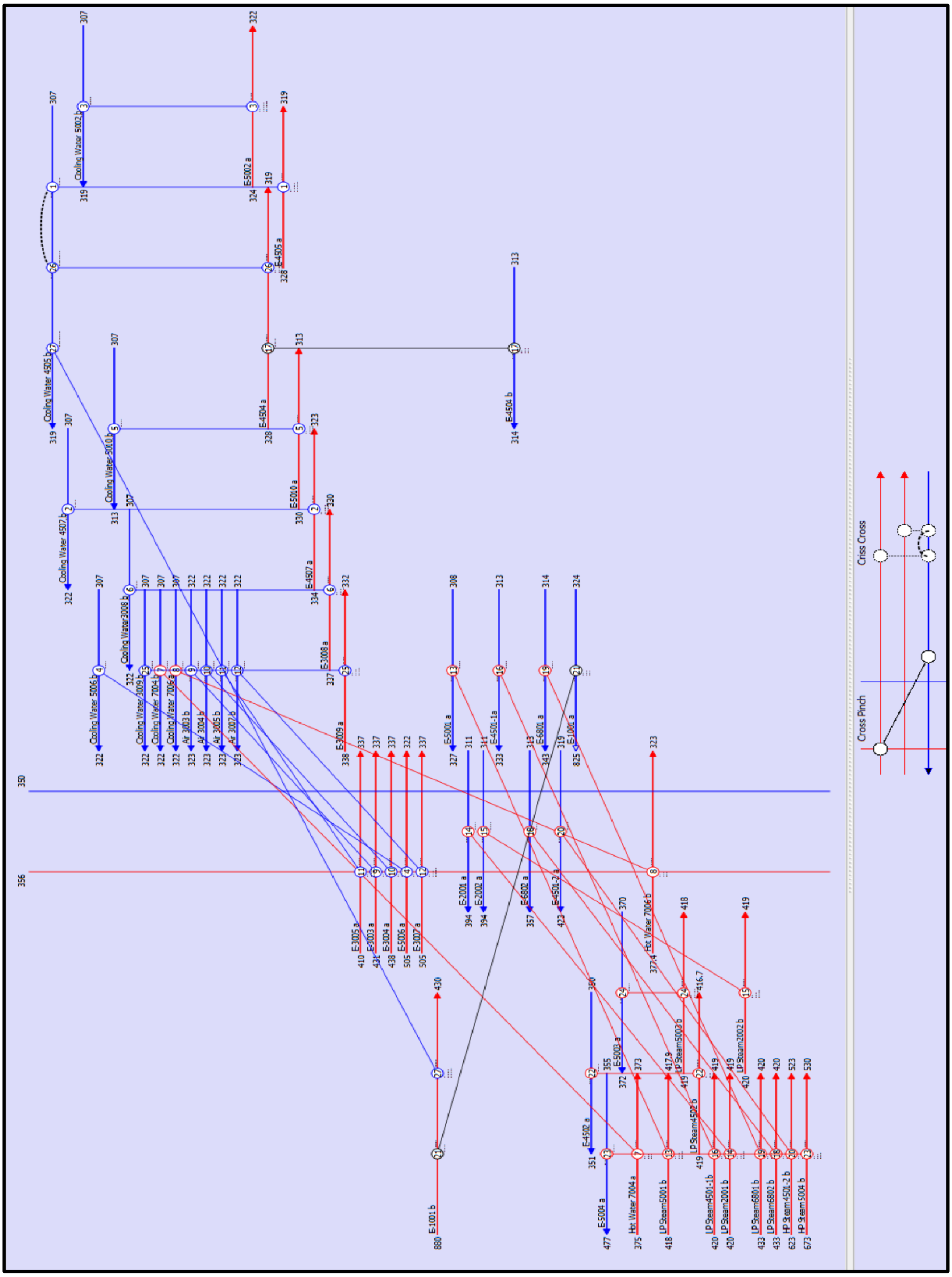

Figure 4.11: Pinch configuration for fixed structure

Table 4.9: The heat exchanger crosses the pinch for fixed structure.

Serial Number	Name	Duty [kW]	Cross pinch duty[kW]	Hot side (Pinch Temperature:356)			
				Stream Name	F*Cp [kW/K]	Inlet Temp [K]	Outlet Temp[K]
27	NewHX1080	74729.7	74729.7	E-1001 b	166.067	879.998	430
21	EX1074	0.28	0.28	E-1001 b	166.067	880	879.998
20	EX1073	110	32.78	HP Steam 4501-2 B	1.1	623	523
19	EX1072	700	700	LP Steam 6801 B	53.84	433	420
18	EX1071	271.3	228.13	LP Steam 6802 B	20.86	433	420
16	EX1069	2500	2500	LP Steam 4501-1B	2500	420	419
15	EX1068	190	89.2771	LP Steam 2002 B	190	420	419
14	EX1067	140	65.7831	LP Steam 2001 B	140	420	419
13	EX1066	600	600	LP Steam 5001 B	6000	418	417.9
12	EX1065	2430	2155.18	E-3007 a	14.46	505	337
11	EX1064	9040	6687.12	E-3005 a	123.83	410	337
10	EX1063	12550	10189.1	E-3004 a	124.25	438	337
9	EX1062	11430	9119.68	E-3003 a	121.59	431	337
8	EX1061	0.03	0.013	Hot Water 7006 A	0.00064	377.4	323
7	EX1060	0.03	0.03	Hot Water 7004 A	0.015	375	373
4	EX1057	1280	1042.19	E-5006 a	6.99454	505	322

Table 4.10: The stream Criss cross heat transfer

Stream	Previous Exchanger				Next Exchanger			
	Hot Stream	F*Cp [kW/K]	Exchanger	Inlet Temp[K]	Hot Stream	F*Cp [kW/K]	Exchanger	Inlet Temp[K]
Cooling Water 4505 B	E-4505 a	1797.78	EX1054	328	E-4504 a	6262.22	NewHX1079	328

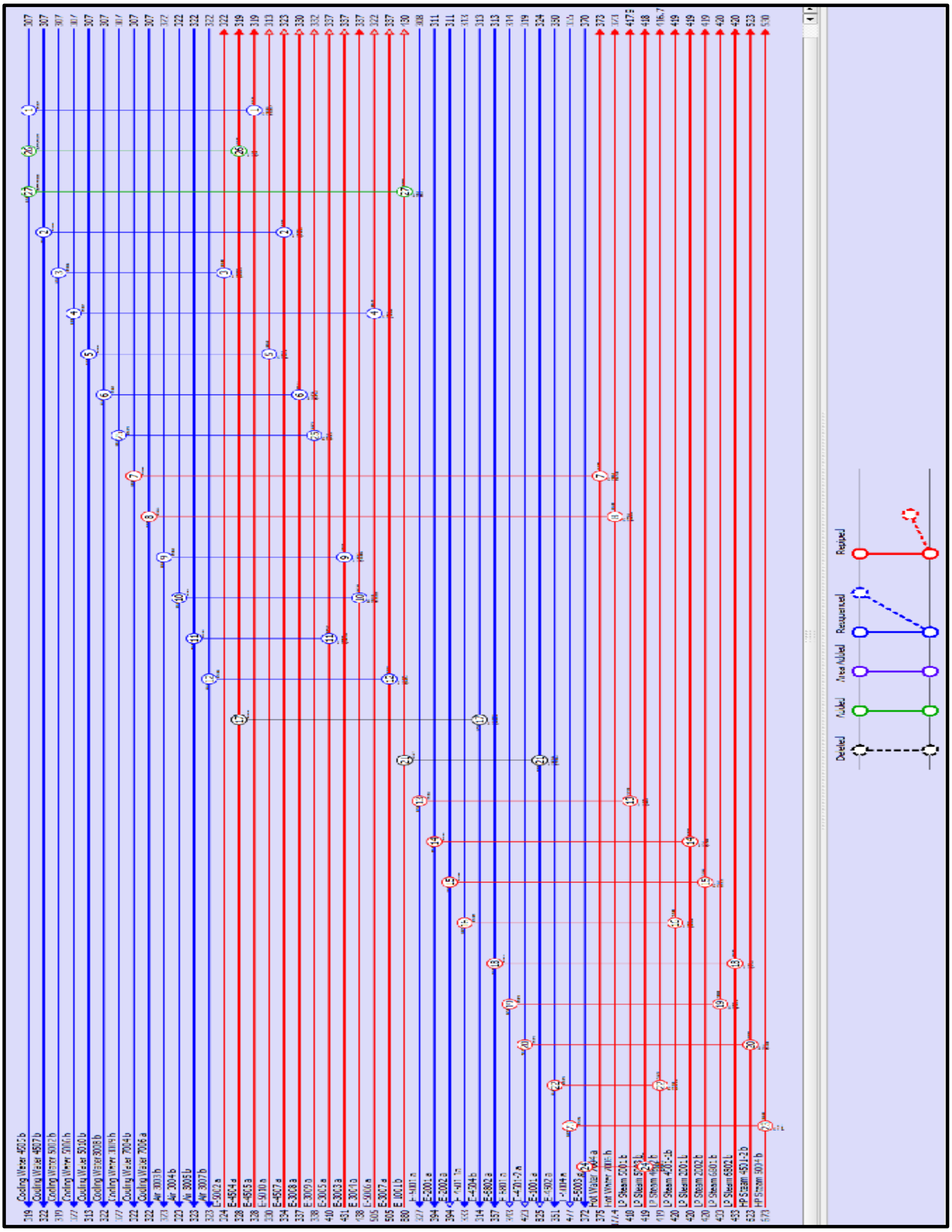

Figure 4.12: Superstructure show the comparing between the existing and fixed structure design

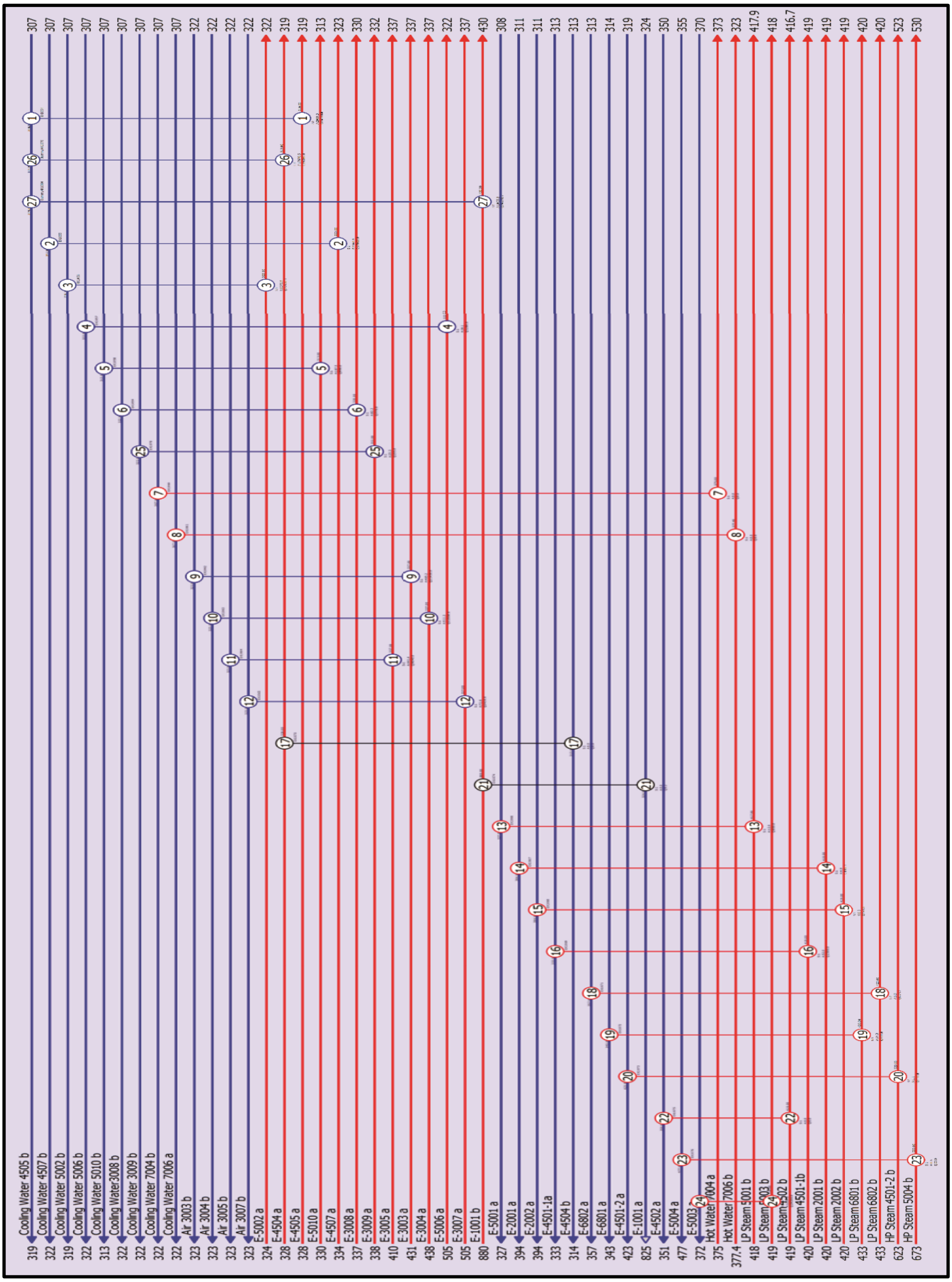

Figure 4.13: Final design for fixed structure

Table 4.11: Summary of result for fixed structure (comparison between the Existing and retrofitting network)

Items		Retrofitting network	Existing Network
ΔTmin	K	15.00	6.00
Hot utility duty	MW	33.761	72.538
Cold utility duty	MW	231.445	128.527
Number of heat exchangers		27.00	25.00
Added area	m^2	32611.91	36236.32
Used area	m^2	32611.91	36236.32
Total annualized cost	MM$/y	130.00	141.07
Total capital cost	MM$/y	9.34	9.18
Total operating cost	MM$/y	120.66	131.89
Hot utility operating cost	MM$/y	15.75	32.94
Cold utility operating cost	MM$/y	84.81	76.97
Other operating cost	MM$/y	20.11	21.98

4.8 Conclusions

The proposed design method is energy efficient and can be used to find the optimum HEN. A reduction of the annual operating cost was obtained by optimizing the HEN for a propene dehydrogenation unit (PDH) using the pinch analysis technique. A revamp design was suggested based on the Simulated Annealing (SA) and the fixed structure techniques. The results of SA optimization were better than the fixed structure. The suggested new design has a saving of 20.65 million US $/year (save up to 14.64%).

Abbreviations

Symbols	Meaning (Units)
A	Area of heat exchanger
a & b	Denote the process stream and utility stream in HENs
CP	Heat capacity flow rate (MW/K)
Cp	Specific heat capacity (Kj/Kg.K)
F*Cp	Heat capacity flow rate (MW/K)
ΔH	Exchanger duty (MW)
HEN	Heat exchanger network
HENs	Heat exchanger networks
HEs	Heat exchangers
MM	Millions
Q	Duty
R	Interest rate (%)
T^T	Outlet temperature of stream (K)
T^S	Inlet temperature of stream (K)
U	Overall heat transfer coefficient (MW/m2 K)
N_{HX}	Total number of heat exchangers
C_{CU}	Cost per unit of cold utility
C_{HU}	Cost per unit of hot utility
C_k	Exponent for area cost

Indices

i	hot process or utility stream

j cold process or utility stream

K index for a heat exchanger, 1 ……. N_{HX}

Binary variables

ψ_{ijk} indicating the existing of match ij at the heat exchanger, k, in optimal network

$\psi_{i,CU}$ indicating the existing of the match between hot stream i and cold utility

$\psi_{\mathrm{HU,j}}$ indicating the existing of the match between hot Utility and cold stream j

Appendix A

Summary for stream after Simulation Annealing [SA]

No	Name	Mass Flow	T^S	T^T	ΔH
		kg/h	K	K	kW
2	E-4504 a	660913.00	328.00	319.00	56360.00
4	E-4505 a	189761.00	328.00	319.00	16180.00
5	E-4507 a	303802.00	334.00	323.00	17800.00
7	E-5002 a	343793.00	324.00	322.00	27630.00
10	E-5006 a	6414.00	505.00	322.00	1280.00
11	E-5010 a	64142.00	330.00	313.00	990.00
13	E-1001 b	175623.00	880.00	430.00	74730.00
14	E-3008 a	176442.00	337.00	330.00	770.00
15	E-3009 a	11840.00	338.00	332.00	255.00
18	E-3003 a	175912.00	431.00	337.00	11430.00
19	E-3004 a	176442.00	438.00	337.00	12550.00
20	E-3005 a	176442.00	410.00	337.00	9040.00
21	E-3007 a	11840.00	505.00	337.00	2430.00
1	E-4501-1a	166321.00	313.00	333.00	2500.00
3	E-4504 b	2207244.00	313.00	314.00	56360.00
6	E-5001 a	6414.00	308.00	327.00	600.00
8	E-5003 a	1954991.00	370.00	372.00	29200.00
9	E-5004 a	201.00	355.00	477.00	50.00

12	E-1001 a	174999.00	324.00	825.00	74730.00
16	E-2001 a	429.00	311.00	394.00	140.00
17	E-2002 a	967.00	311.00	394.00	190.00
22	E-4502 a	614811.00	350.00	351.00	24210.00
23	E-4501-2 a	250.00	319.00	423.00	110.00
24	E-6801 a	25560.00	314.00	343.00	700.00
25	E-6802 a	6000.00	313.00	357.00	271.30

Appendix B

Utility after Simulation Annealing [SA]

No	Name	T^S	T^T	Duty	Operation Cost Index	Annual Operation Cost	Capital Cost
		K	K	kW	M$/Kj	M$	0.3985
1	LP Steam 4501-1b	420.00	419.00	1586.44	1.505E-05	0.7394	0.2910
4	LP Steam 5001 b	418.00	417.90	1025.62	1.505E-05	0.4780	3.7860
6	LP Steam 5003 b	419.00	418.00	29197.25	1.505E-05	13.6078	0.0575
7	HP Steam 5004 b	673.00	530.00	53.15	1.720E-05	0.0283	0.0858
12	LP Steam 2001 b	420.00	419.00	140.00	1.505E-05	0.0652	0.0865
13	LP Steam 2002 b	420.00	419.00	142.54	1.505E-05	0.0664	3.2627
18	LP Steam 4502 b	419.00	416.70	24199.51	1.505E-05	11.2785	0.0766
19	HP Steam 4501-2 b	623.00	523.00	110.00	1.720E-05	0.0586	0.2254
20	LP Steam 6801 b	433.00	420.00	710.47	1.505E-05	0.3311	0.1223
21	LP Steam 6802 b	433.00	420.00	271.30	1.505E-05	0.1264	0.0334
22	Hot Water 7004 a	375.00	373.00	0.00	1.290E-05	0.0000	0.0334
24	Hot Water 7006 b	377.40	323.00	0.00	1.290E-05	0.0000	2.3616
2	Cooling Water 4505 b	307.00	319.00	16076.71	3.989E-06	1.9856	2.5592
3	Cooling Water 4507 b	307.00	322.00	17800.06	3.190E-06	1.7579	3.6240
5	Cooling Water 5002 b	307.00	319.00	27630.00	3.989E-06	3.4125	0.3972
8	Cooling Water 5006 b	307.00	322.00	1579.56	3.190E-06	0.1560	0.1259
9	Cooling Water 5010 b	307.00	313.00	285.05	7.974E-06	0.0704	0.3447
10	Cooling Water3008 b	307.00	322.00	1299.80	3.190E-06	0.1284	0.1180
11	Cooling Water 3009 b	307.00	322.00	255.00	3.190E-06	0.0252	1.7525
14	Air 3003 b	322.00	323.00	11003.53	5.556E-05	18.9261	1.9432
15	Air 3004 b	322.00	323.00	12550.03	5.556E-05	21.5860	1.4856
16	Air 3005 b	322.00	323.00	8911.66	5.556E-05	15.3281	0.5466
17	Air 3007 b	322.00	323.00	2428.32	5.556E-05	4.1767	0.0334
23	Cooling Water 7004 b	307.00	322.00	0.00	3.190E-06	0.0000	0.0334
25	Cooling Water 7006 a	307.00	322.00	0.00	3.190E-06	0.0000	0.3985

Appendix(C)

Stream for fixed structure after retrofitting

No	Name	Mass Flow	T^S	T^T	ΔH
		kg/h	K	K	kW
2	E-4504 a	660913	328	319	56360.0
4	E-4505 a	189761	328	319	16180.0
5	E-4507 a	303802	334	323	17800.0
7	E-5002 a	343793	324	322	27630.0
10	E-5006 a	6414	505	322	1280.0
11	E-5010 a	64142	330	313	990.0
13	E-1001 b	175623	880	430	74730.0
14	E-3008 a	176442	337	330	770.0
15	E-3009 a	11840	338	332	255.0
18	E-3003 a	175912	431	337	11430.0
19	E-3004 a	176442	438	337	12550.0
20	E-3005 a	176442	410	337	9040.0
21	E-3007 a	11840	505	337	2430.0
1	E-4501-1a	166321	313	333	2500.0
3	E-4504 b	2207244	313	314	56360.0
6	E-5001 a	6414	308	327	600.0
8	E-5003 a	1954991	370	372	29200.0
9	E-5004 a	201	355	477	50.0

12	E-1001 a	174999	324	825	74730.0
16	E-2001 a	429	311	394	140.0
17	E-2002 a	967	311	394	190.0
22	E-4502 a	614811	350	351	24210.0
23	E-4501-2 a	250	319	423	110.0
24	E-6801 a	25560	314	343	700.0
25	E-6802 a	6000	313	357	271.3

Appendix (D)

Utility after fixed structure

No	Name	T^S	T^T	Duty	Operation Cost Index	Annual Operation Cost	Capital Cost
		K	K	kW	$/kJ	M$/y	M$
1	LP Steam 4501-1b	420	419	2500	1.505E-05	1.1652	0.5587
4	LP Steam 5001 b	418	418	600	1.505E-05	0.2796	0.2011
6	LP Steam 5003 b	419	418	29200	1.505E-05	13.6091	3.7863
7	HP Steam 5004 b	673	530	50	1.720E-05	0.0266	0.0564
12	LP Steam 2001 b	420	419	140	1.505E-05	0.0652	0.0858
13	LP Steam 2002 b	420	419	190	1.505E-05	0.0886	0.1003
18	LP Steam 4502 b	419	417	0	1.505E-05	0.0000	0.0334
19	HP Steam 4501-2 b	623	523	110	1.720E-05	0.0586	0.0766
20	LP Steam 6801 b	433	420	700	1.505E-05	0.3262	0.2232
21	LP Steam 6802 b	433	420	271	1.505E-05	0.1264	0.1223
22	Hot Water 7004 a	375	373	0	1.290E-05	0.0000	0.0334
24	Hot Water 7006 b	377	323	0	1.290E-05	0.0000	0.0334
2	Cooling Water 4505 b	307	319	147270	3.989E-06	18.1891	13.7281
3	Cooling Water 4507 b	307	322	17800	3.190E-06	1.7579	2.5592
5	Cooling Water 5002 b	307	319	27630	3.989E-06	3.4125	3.6240
8	Cooling Water 5006 b	307	322	1280	3.190E-06	0.1264	0.3409
9	Cooling Water 5010 b	307	313	990	7.974E-06	0.2444	0.2838
10	Cooling Water3008 b	307	322	770	3.190E-06	0.0760	0.2382
11	Cooling Water 3009 b	307	322	255	3.190E-06	0.0252	0.1180
14	Air 3003 b	322	323	11430	5.556E-05	19.6596	1.8056
15	Air 3004 b	322	323	12550	5.556E-05	21.5860	1.9432
16	Air 3005 b	322	323	9040	5.556E-05	15.5488	1.5023
17	Air 3007 b	322	323	2430	5.556E-05	4.1796	0.5469
23	Cooling Water 7004 b	307	322	0	3.190E-06	0.0000	0.0334
25	Cooling Water 7006 a	307	322	0	3.190E-06	0.0000	0.0334

Appendix E

Matching after Simulation Annealing [SA]

No	Tag	Stream	Mass Flow	Inlet T	Outlet T	LMTD	Q	A	Added A	Capital Cost
			kg/h	K	K	K	kW	m²	m²	M$
7	EX1060	E-5002 a	343793.00	324.00	324.00	13.00	0.00	0.0	0.0	0.0000
		E-2002 a	967.00	311.00	311.00					
8	EX1061	E-5002 a	343793.00	324.00	324.00	11.00	0.00	0.0	0.0	0.0000
		E-4501-1a	166321.00	313.00	313.00					
17	EX1070	E-4504 a	660913.00	328.00	319.00	9.44	56360.0	11938.4	11938.4	1.4610
		E-4504 b	2207244.00	313.00	314.00					
21	EX1074	E-1001 b	175623.00	880.00	430.00	77.73	74729.6	1922.8	1922.8	0.3390
		E-1001 a	174999.00	324.00	825.00					
27	EX23740	E-3005 a	176442.00	410.00	409.63	88.46	45.77	2.1	2.1	0.0014
		E-2002 a	967.00	311.00	330.99					
29	EX16984	E-5010 a	64142.00	330.00	326.25	12.07	218.14	72.3	72.3	0.0246
		E-4501-1a	166321.00	315.16	316.90					
36	EX8580	E-5010 a	64142.00	326.25	321.62	9.81	269.78	110.1	110.1	0.0344
		E-4501-1a	166321.00	313.00	315.16					
38	EX33749	E-3007 a	11840.00	505.00	504.88	111.31	1.68	0.1	0.1	0.0001
		E-2002 a	967.00	393.26	394.00					
1	EX1054	E-4505 a	189761.00	327.94	319.00	10.40	16076.71	3092.7	3092.7	0.4959
		Cooling Water 4505 b	1154427.37	307.00	319.00					

2	EX1055	E-4507 a	303802.00	334.00	323.00	13.90	17800.06	2560.4	2560.4	0.4263
		Cooling Water 4507 b	1022013.15	307.00	322.00					
3	EX1056	E-5002 a	343793.00	324.00	322.00	9.10	27630.00	6070.9	6070.9	0.8505
		Cooling Water 5002 b	1984039.70	307.00	319.00					
4	EX1057	E-5006 a	6414.00	505.00	322.00	63.01	1279.98	40.6	40.6	0.0155
		Cooling Water 5006 b	90692.16	309.84	322.00					
5	EX1058	E-5010 a	64142.00	316.48	313.00	5.60	202.50	72.3	72.3	0.0246
		Cooling Water 5010 b	40916.64	307.00	311.26					
6	EX1059	E-3008 a	176442.00	337.00	330.00	20.85	770.04	73.9	73.9	0.0250
		Cooling Water3008 b	74629.70	308.19	317.08					
9	EX1062	E-3003 a	175912.00	427.49	337.00	46.10	11003.53	477.3	477.3	0.1112
		Air 3003 b	39612711.41	322.00	323.00					
10	EX1063	E-3004 a	176442.00	438.00	337.00	49.09	12550.03	511.3	511.3	0.1175
		Air 3004 b	45180095.59	322.00	323.00					
11	EX1064	E-3005 a	176442.00	408.96	337.00	40.65	8911.66	438.5	438.5	0.1039
		Air 3005 b	32081970.24	322.00	323.00					
12	EX1065	E-3007 a	11840.00	504.88	337.00	66.88	2428.32	72.6	72.6	0.0247
		Air 3007 b	8741940.76	322.00	323.00					

13	EX1066	LP Steam 5001 b	8790994.20	417.96	417.90	100.13	600.01	12.0	12.0	0.0058
		E-5001 a	6414.00	308.00	327.00					
14	EX1067	LP Steam 2001 b	119998.60	420.00	419.00	57.58	140.00	4.9	4.9	0.0028
		E-2001 a	429.00	311.00	394.00					
15	EX1068	LP Steam 2002 b	122180.99	420.00	419.00	51.43	142.54	5.5	5.5	0.0031
		E-2002 a	967.00	330.99	393.26					
16	EX1069	LP Steam 4501-1b	2598125.54	420.00	419.00	92.72	1586.44	34.2	34.2	0.0135
		E-4501-1a	166321.00	320.31	333.00					
18	EX1071	LP Steam 6802 b	34197.35	433.00	420.00	90.62	271.30	6.0	6.0	0.0033
		E-6802 a	6000.00	313.00	357.00					
19	EX1072	LP Steam 6801 b	46844.00	432.81	420.00	97.68	700.00	14.3	14.3	0.0067
		E-6801 a	25560.00	314.00	343.00					
20	EX1073	HP Steam 4501-2 b	197.45	623.00	523.00	201.99	110.00	1.1	1.1	0.0009
		E-4501-2 a	250.00	319.00	423.00					
22	EX1075	LP Steam 4502 b	9018451.49	419.00	416.70	67.35	24199.51	718.6	718.6	0.1543

		E-4502 a	614811.00	350.00	351.00					
23	EX1076	HP Steam 5004 b	358.75	673.00	538.48	189.67	50.00	0.5	0.5	0.0005
		E-5004 a	201.00	355.00	477.00					
24	EX1077	LP Steam 5003 b	25026214.63	419.00	418.00	47.50	29197.25	1229.4	1229.4	0.2371
		E-5003 a	1954991.00	370.00	372.00					
25	EX1078	E-3009 a	11840.00	338.00	332.00	20.17	255.00	25.3	25.3	0.0106
		Cooling Water 3009 b	14641.24	307.00	322.00					
26	EX11265	E-5010 a	64142.00	321.62	316.48	10.59	299.58	113.2	113.2	0.0352
		Cooling Water 5006 b	90692.16	307.00	309.84					
28	EX30713	LP Steam 6801 b	46844.00	433.00	432.81	81.90	10.46	0.5	0.5	0.0005
		E-4502 a	614811.00	351.00	351.00					
30	EX14587	LP Steam 5001 b	8790994.20	418.00	417.96	99.36	425.61	17.1	17.1	0.0078
		E-4501-1a	166321.00	316.90	320.31					
31	EX16551	E-3003 a	175912.00	429.58	427.49	109.99	253.71	9.2	9.2	0.0047
		Cooling Water3008 b	74629.70	317.08	320.01					

32	EX40014	HP Steam 5004 b	358.75	531.08	530.00	206.54	0.40	0.0	0.0	0.0000
		E-1001 a	174999.00	324.00	324.00					
33	EX35778	HP Steam 5004 b	358.75	538.48	531.08	162.75	2.75	0.1	0.1	0.0001
		E-5003 a	1954991.00	372.00	372.00					
34	EX15482	E-4505 a	189761.00	328.00	327.94	20.37	103.28	20.3	20.3	0.0089
		Cooling Water3008 b	74629.70	307.00	308.19					
35	EX21080	E-3005 a	176442.00	409.63	408.96	97.16	82.55	3.4	3.4	0.0021
		Cooling Water 5010 b	40916.64	311.26	313.00					
37	EX12073	E-3003 a	175912.00	431.00	429.58	109.29	172.77	6.3	6.3	0.0035
		Cooling Water3008 b	74629.70	320.01	322.00					

Appendix F

Matching after Fixed Structure

No	Tag	Stream	Mass Flow	Inlet T	Outlet T	LMTD	Q	A	Added A	Capital Cost
			kg/h	K	K	K	kW	m²	m²	M$
17	EX1070	E-4504 a	6609E3:E4713	328.00	328.00	15.00	0.01	0.0	0.0	0.0000
		E-4504 b	2207244.00	313.00	313.00					
21	EX1074	E-1001 b	175623.00	880.00	880.00	556.00	0.29	0.0	0.0	0.0000
		E-1001 a	174999.00	324.00	324.00					
1	EX1054	E-4505 a	189761.00	328.00	319.00	15.53	16180.00	2084.3	2084.3	0.3616
		Cooling Water 4505 b	10575061.37	307.00	308.32					
2	EX1055	E-4507 a	303802.00	334.00	323.00	13.90	17800.00	2560.4	2560.4	0.4263
		Cooling Water 4507 b	1022009.57	307.00	322.00					
3	EX1056	E-5002 a	343793.00	324.00	322.00	9.10	27630.00	6070.9	6070.9	0.8505
		Cooling Water 5002 b	1984039.70	307.00	319.00					
4	EX1057	E-5006 a	6414.00	505.00	322.00	67.16	1280.00	38.1	38.1	0.0147
		Cooling Water 5006 b	73492.82	307.00	322.00					
5	EX1058	E-5010 a	64142.00	330.00	313.00	10.56	990.00	187.5	187.5	0.0527

		Cooling Water 5010 b	142105.26	307.00	313.00					
6	EX1059	E-3008 a	176442.00	337.00	330.00	18.72	770.00	82.3	82.3	0.0272
		Cooling Water3008 b	44210.53	307.00	322.00					
7	EX1060	Hot Water 7004 a	29.10	375.00	375.00	68.00	0.00	0.0	0.0	0.0000
		Cooling Water 7004 b	1.77	307.00	307.00					
8	EX1061	Hot Water 7006 b	0.05	377.40	377.40	70.40	0.00	0.0	0.0	0.0000
		Cooling Water 7006 a	2.03	307.00	307.00					
9	EX1062	E-3003 a	175912.00	431.00	337.00	47.11	11430.00	485.2	485.2	0.1127
		Air 3003 b	41148000.00	322.00	323.00					
10	EX1063	E-3004 a	176442.00	438.00	337.00	49.09	12550.00	511.3	511.3	0.1175
		Air 3004 b	45180000.00	322.00	323.00					
11	EX1064	E-3005 a	176442.00	410.00	337.00	40.96	9040.00	441.4	441.4	0.1045
		Air 3005 b	32544000.00	322.00	323.00					
12	EX1065	E-3007 a	11840.00	505.00	337.00	66.91	2430.00	72.6	72.6	0.0247
		Air 3007 b	8748000.00	322.00	323.00					

13	EX1066	LP Steam 5001 b	5142857.14	418.00	417.90	100.15	600.00	12.0	12.0	0.0058
		E-5001 a	6414.00	308.00	327.00					
14	EX1067	LP Steam 2001 b	120000.00	420.00	419.00	57.58	140.00	4.9	4.9	0.0028
		E-2001 a	429.00	311.00	394.00					
15	EX1068	LP Steam 2002 b	162857.14	420.00	419.00	57.58	190.00	6.6	6.6	0.0036
		E-2002 a	967.00	311.00	394.00					
16	EX1069	LP Steam 4501-1b	4094266.10	420.00	419.00	96.19	2500.00	52.0	52.0	0.0189
		E-4501-1a	166321.00	313.00	333.00					
18	EX1071	LP Steam 6802 b	34197.64	433.00	420.00	90.62	271.30	6.0	6.0	0.0033
		E-6802 a	6000.00	313.00	357.00					
19	EX1072	LP Steam 6801 b	46153.85	433.00	420.00	97.78	700.00	14.3	14.3	0.0067
		E-6801 a	25560.00	314.00	343.00					
20	EX1073	HP Steam 4501-2 b	197.44	623.00	523.00	201.99	110.00	1.1	1.1	0.0009
		E-4501-2 a	250.00	319.00	423.00					
22	EX1075	LP Steam 4502 b	13.92	419.00	419.00	69.00	0.00	0.0	0.0	0.0000

		E-4502 a	614811.00	350.00	350.00					
23	EX1076	HP Steam 5004 b	337.46	673.00	530.00	185.30	50.00	0.5	0.5	0.0005
		E-5004 a	201.00	355.00	477.00					
24	EX1077	LP Steam 5003 b	25028571.43	419.00	418.00	47.50	29200.00	1229.5	1229.5	0.2371
		E-5003 a	1954991.00	370.00	372.00					
25	EX1078	E-3009 a	11840.00	338.00	332.00	20.17	255.00	25.3	25.3	0.0106
		Cooling Water 3009 b	14641.15	307.00	322.00					
26	New HX1079	E-4504 a	660913.00	328.00	319.00	12.76	56359.99	17670.5	17670.5	1.9994
		Cooling Water 4505 b	10575061.37	308.32	312.91					
27	New HX1080	E-1001 b	175623.00	880.00	430.00	283.33	74729.71	1055.2	1055.2	0.2098
		Cooling Water 4505 b	10575061.37	312.91	319.00					

CHAPTER 5

Retrofitting and analysis of the heat exchanger network for the fractionation unit

5.1 Introduction

Since the start of industrialization and technological developments, the world energy demand has grown exponentially. Fulfilling such a tremendous energy demand has both economical as well as environmental repercussions. Owing to the economic burden and adverse consequences of higher energy consumption, considerable research efforts have been done to minimize the energy the consumption mainly by heat recovery and heat integration in the industries [25]. Petrochemical industries play a vital role in stabilizing the economy of both developed and developing countries and these industries are energy intensive i.e. high energy consumption [3], [26]–[28], [39]. An effective route to maximize the heat recovery, and minimize the energy demand in a process industry is the use of Pinch technology [5], [8], [29]–[31].. Various methodologies have been proposed over time which can be divided into two different design tasks (1) A grass root problem, in which a new heat exchanger network is designed and (2) A retrofit, in which an existing HEN is optimized for energy efficiency [40]. Designing HEN network based on Pinch method was originally proposed by Linnhoff and Hindmarsh [26], [39]. This one of the widely used method in industrial applications, in which the HEN design is finalized after setting the Energy Targets by using Composite Curves or Problem Table Algorithm [39], [41], [42]. Al-Mutairi [19] employed the pinch technique to optimize the HEN of a fluid catalytic cracking (FCC) unit, by means of optimum ΔT_{min}. The proposed design improved the heat recovery and energy cost of the FCC unit. Babaqi et al [43] studied the pinch analysis and retrofit design of HEN for Continuous Catalytic

Regeneration Reformer Process. This retrofitting saving up to 32%.it was done through adding three HE with shell tube of two heat exchangers.

Recently, an upsurge of interest has been observed, in employing the pinch concepts in formulating mathematical models for reducing the total annual operating cost [1], [20]. Floudas and Grossmann [10], [32] proposed that mixed-integer optimization can be utilized to handle the problems of flexibility analysis. The authors utilized linear programming to minimize the utility cost and determine the pinch point for the given constraints and parameters. Furthermore, the minimum numbers of units were determined by Mixed Integer Linear Programming (MILP).

However, it should be noted that decomposing the Heat Exchanger Network (HEN) problem into separate targeting procedure (i.e., utility, energy, and area) does not guarantee that the total costs are minimized [15], [33], [34]. Mixed Integer Nonlinear Programming (MINLP) formulation is efficient enough to capture the common features of the HEN and guarantee minimization of the total/overall cost of the network[7], [9], [12].

In this paper, an improved heat exchanger networks (HENs) for optimum performance of the Fractionation unit at a minimum Total Annual Cost is developed. Two techniques were adopted to optimize and retrofit the HEN, (1) the Simulated Annealing (SA) and (2) fixed structure technique using MINLP formulation. In the simulated annealing technique, the HEN is flexible and new heat exchanger can be added or removed. While in the fixed structure technique, the system allows a very limited space for a change of the number of heat exchangers in the network. The results of optimization can be applied on the Fractionation unit by selecting the feasible revamp scheme. The results of our improved HENs will be compared to those of existing HEN.

Process Description

Fractionation section consist of three main parts: (a) Deethanizer section (Unit 4500): This section is designed to remove ethane and lighter materials received from the separation system to meet the propylene product purity specifications, (b) Refrigeration system (ME-4501): The main purpose of the refrigeration unit is to condense the propylene vapors received from deethanizer section and to separate oil from propylene vapors, and (c) Propylene-Propane Splitter Section (Unit 4500): This section receives product from the Deethanizer Stripper and separates propane from unconverted propane. Figure (5.1) shows the simplified process flow diagram of fractionation section.

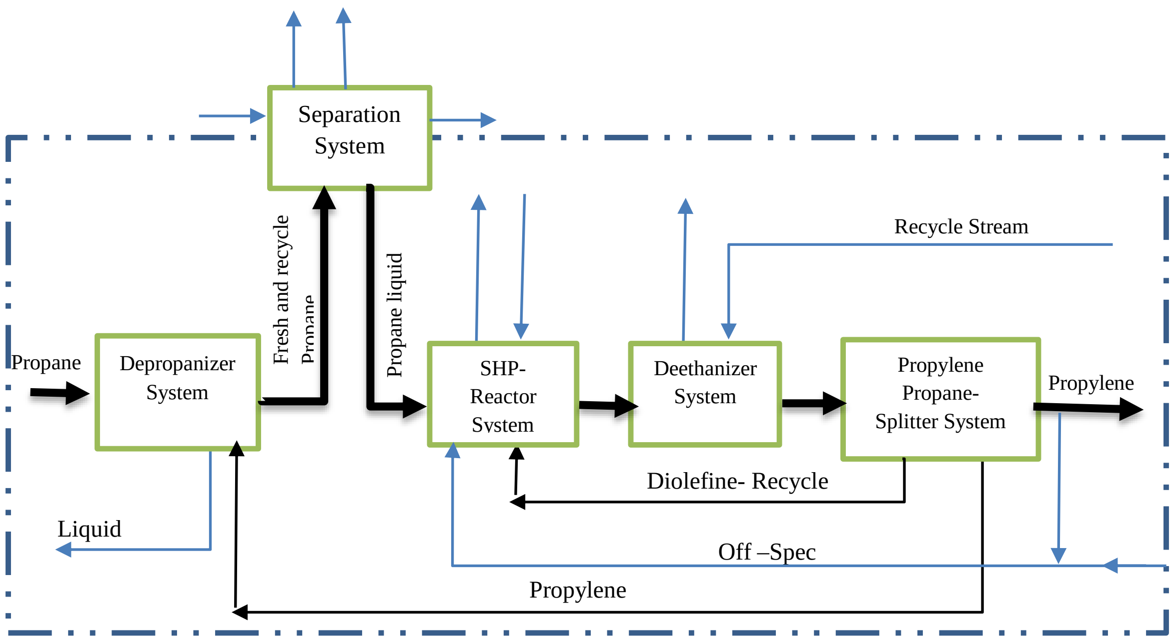

Figure 5.1: the simplified process flow diagram of fractionation section

5.2 Existing heat exchanger network

The current process consists of 12 heat exchangers. There is a total of 24 streams. 7 cold process streams and 6 hot process streams as well as 6 hot and 5 cold utilities streams.

The hot exchanger's utility uses either HP or LP steam the cold utilities heat exchangers use is cooling water. The details of all the streams and utilities are listed in Table (5.1) and (5.2), respectively. The grid diagram of existing heat exchanger network of fractionation unit is shown in figure (5.2).

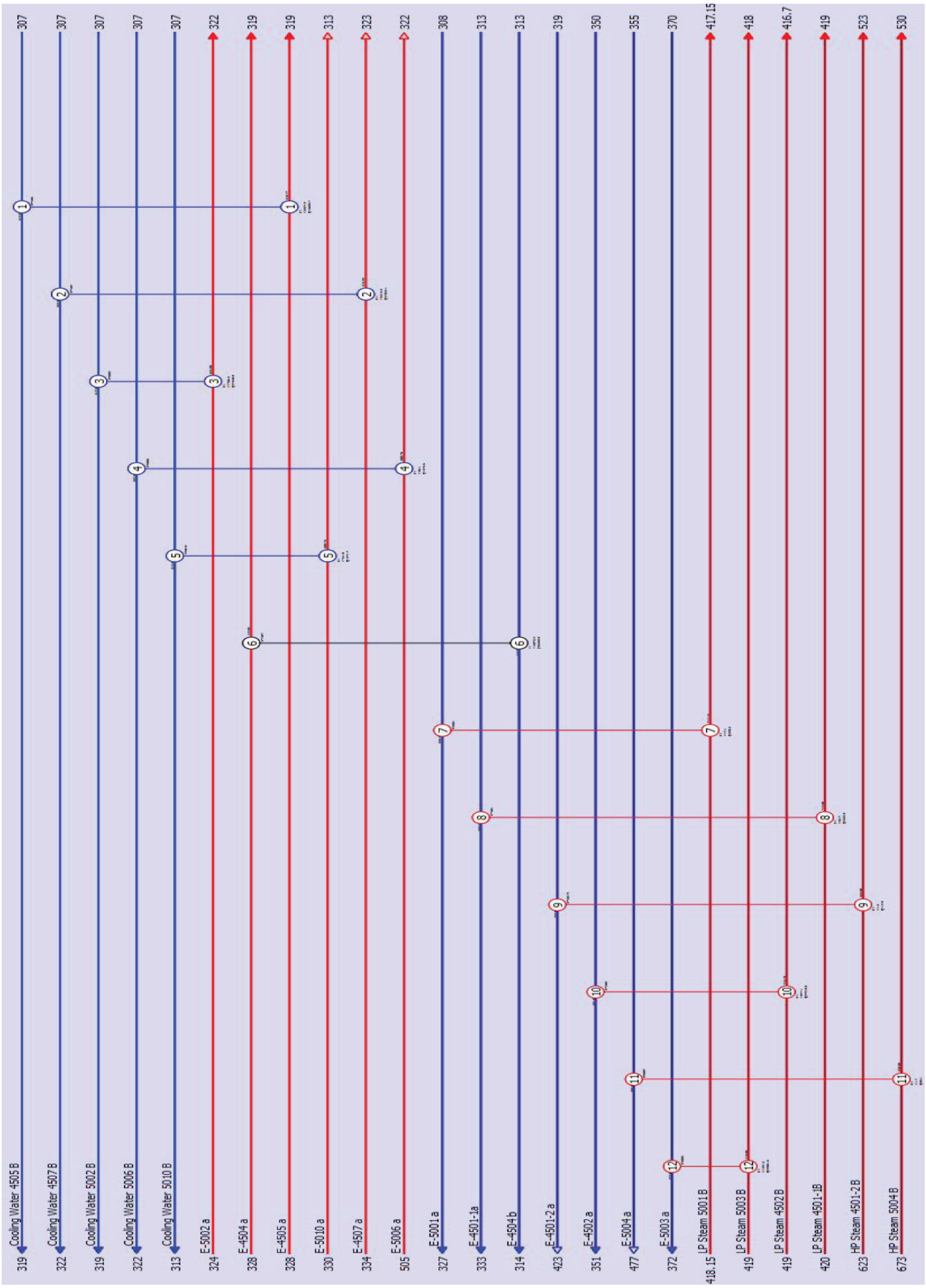

Figure 5.2: Grid diagram of existing heat exchanger network of the Fractionation Unit

Table 5.1: The hot and cold stream for existing design.

No.	Stream	Flow rate [kg/h]	TS [K]	TT [K]	MCp [kW/K]	DH [kW]	FR [(m²•K)/ kW]	Inlet Pressure [bar]	Cp [kJ/(kg•K)]
1	E-4501-1a	166321	313	333	126.13	2522.54	0.00026	37	2.73
2	E-4504 a	660913	328	319	6297.03	56673.3	0.0001	18.5	34.3
3	E-4504 b	2207244	313	314	61269.4	61269.4	0.0001	13.2	99.93
4	E-4505 a	189761	328	319	1799.04	16191.4	0.00026	18.2	34.13
5	E-4507 a	303802	334	323	1628.72	17915.9	0.00026	29.3	19.3
6	E-5001 a	6414	308	327	31.59	600.19	0.00026	18.4	17.73
7	E-5002 a	343793	324	322	13816.7	27633.3	0.00026	16.9	144.68
8	E-5003 a	1954991	370	372	14602.7	29205.4	0.00035	17.8	26.89
9	E-5004 a	201	355	477	0.42	50.07	0.00026	4	7.35
10	E-5006 a	6414	505	322	7.00	1281.36	0.00026	17.4	3.93
11	E-5010 a	64142	330	313	58.44	993.49	0.00026	24.1	3.28
12	E-4502 a	614811	350	351	24213.3	24213.3	0.00035	30.3	141.78
13	E-4501-2 a	250	319	423	1.06	110	0.00026	37	15.23

Table 5.2: The hot and cold utilities for existing design

	Utility	T^S [K]	T^T [K]	FR [(m²•K)/ KW]	Flow rate [kg/h]	Cp [kJ/ (kg•K)]	Inlet Pressure [bar]
1	LP Steam 4501-1B	420	419	9.00E-05	3979588.57	2.31	4
2	Cooling Water 4505 B	307	319.0	0.0002	1192389.12	4.18	4.5
3	Cooling Water 4507 B	307	322.0	0.0002	1598233.31	4.18	4.5
4	LP Steam 5001 B	418.2	417.2	0.0001	925025.57	2.33	4
5	Cooling Water 5002 B	307	319.0	0.0002	1984039.70	4.18	4.5
6	LP Steam 5003 B	419	418.0	0.0002	48781053.9	2.21	3.2
7	HP Steam 5004 B	673	530.0	0.0001	820.83	2.61	36
8	Cooling Water 5006 B	307	322.0	0.0002	79077.57	4.18	4.5
9	Cooling Water 5010 B	307	313.0	0.0002	178287.32	4.18	4.5
10	LP Steam 4502 B	419	416.7	0.0001	17165316.63	2.21	3.2
11	HP Steam 4501-2 B	623	523.0	0.0001	1446.69	2.79	36

5.3 Cost Data and Operation Cost

In order to take a decision about the economic feasibility of the revamp plant, an economical evaluation must be performed. The typical operation time is 8600 hours, while the plant is assumed to have a lifetime of 5 years. The annual interest rate is assigned to be roughly 6%. The mathematical formula, used to determine the annualization factor [38], is given by:

$$\text{Annualization factor} = R\,(1+R)^{n} / (1+R)^{n} - 1$$

The operation costs are mainly related to the consumption of fuel for heat generation. This heat is necessary for steam generation in order to supply the hot streams. Moreover, the cost associated with cold utilities is also included, but it is in lower side in comparison with that required for the hot utilities. Table (5.3) below lists the cost data of the utilities from the local company in Saudi Arabia.

Table 5.3: The fuel price of the hot and cold utilities

Hot Utility	**Fuel price**
HP steam	7.2 US $/ton
LP steam	6.2 US $/ton
Cold Utility	
Cooling water	0.2 US $/ton

Capital cost

Heat exchanger (HE) and utility cost are given by the following equations:

HE cost = $A_1 + B_1 (\text{area})^{C_1}$

Energy cost (Utility) = $A_2 + B_2 (\text{Duty})^{C_2}$

Where A represent a fixed cost.

B_1, the heat exchanger cost per unit, depend on the type of material as indicated in Table (5.4).

Table 5.4: The Price of heat exchanger's material

Material	Density kg/m3	Material price $/t	Thermal conductivity W/(m.k)
CS	7850	530	46.7
SS	7930	550	19

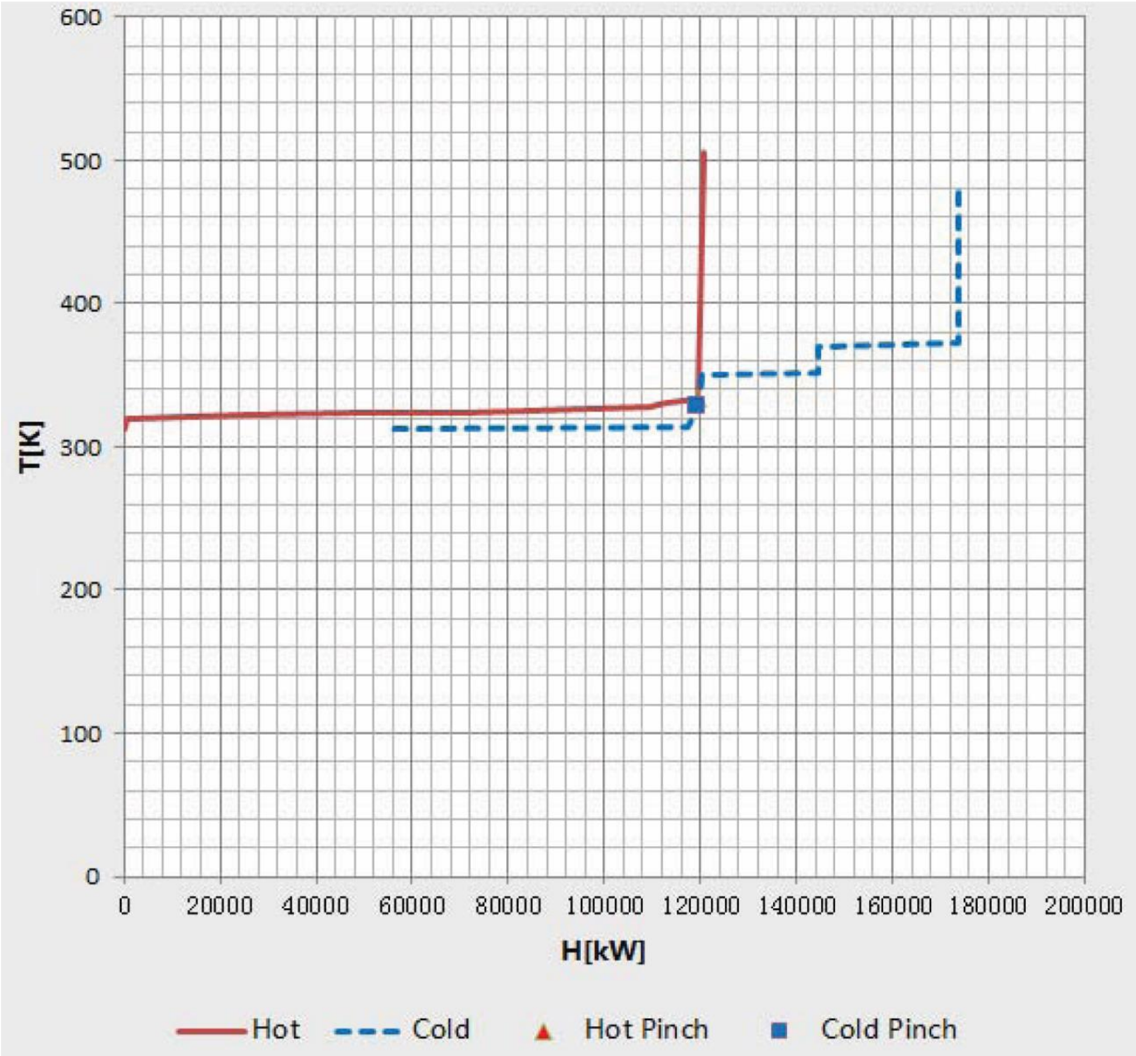

Figure 5.3: Composite Curves for existing Heat Exchanger Network of the Fractionation Unit

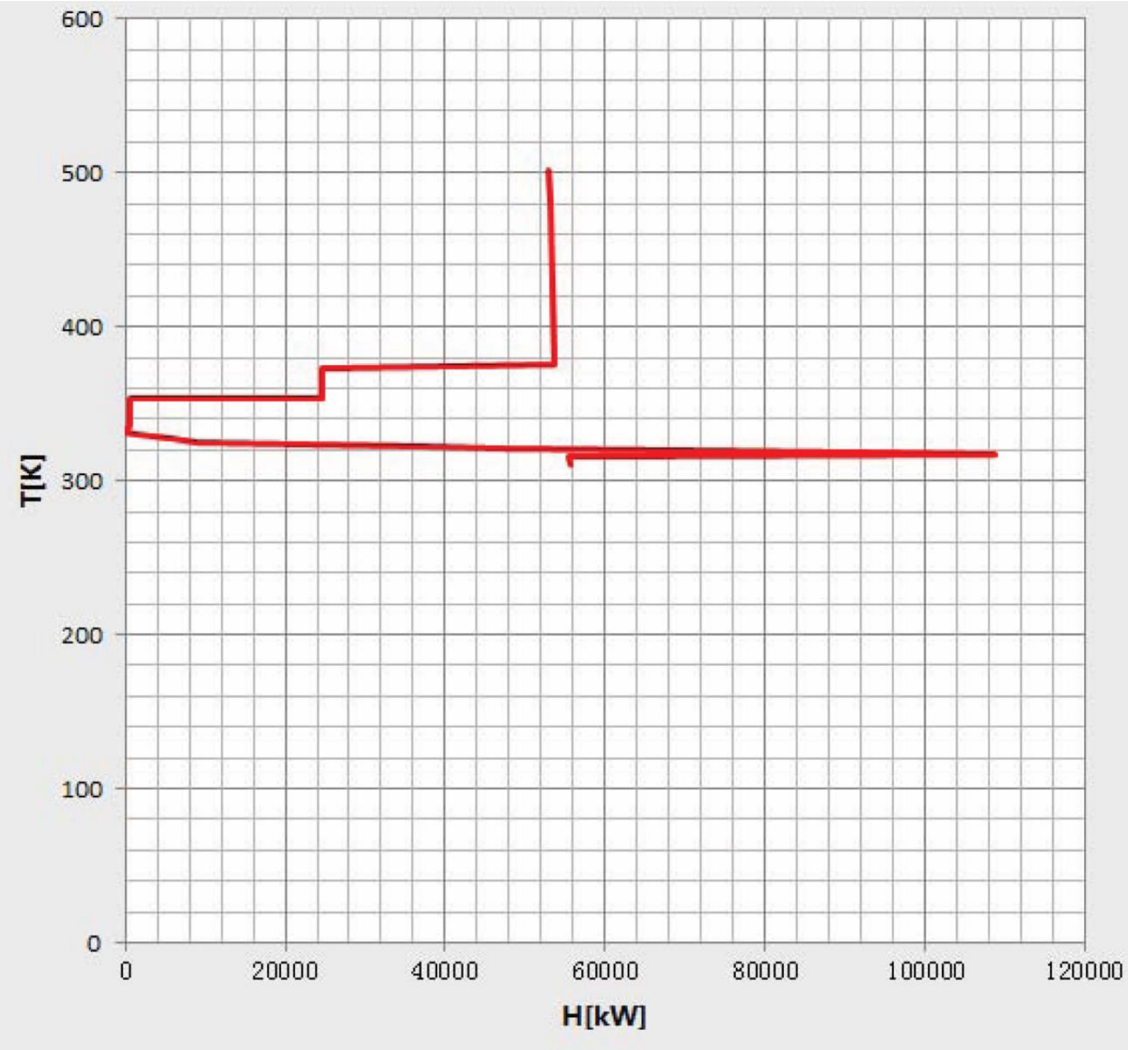

Figure 5.4: The Existing Heat Exchanger Network Grand Composite Curve of the Fractionation Unit

5.4 Result and Discussion

5.4.1 Pinch Analysis

Pinch Analysis is used, and it was found that the hot and cold utility demands of the existing network were 57.42 [MW] and 74.68 [MW], respectively as shown in compost curve figure (5.3). In addition, the grand compost curve (GCC), for the existing plant (Fractionation unit), presented the variation of heat supply and demand as shown in figure (5.4).

5.4.2 Pinch configuration

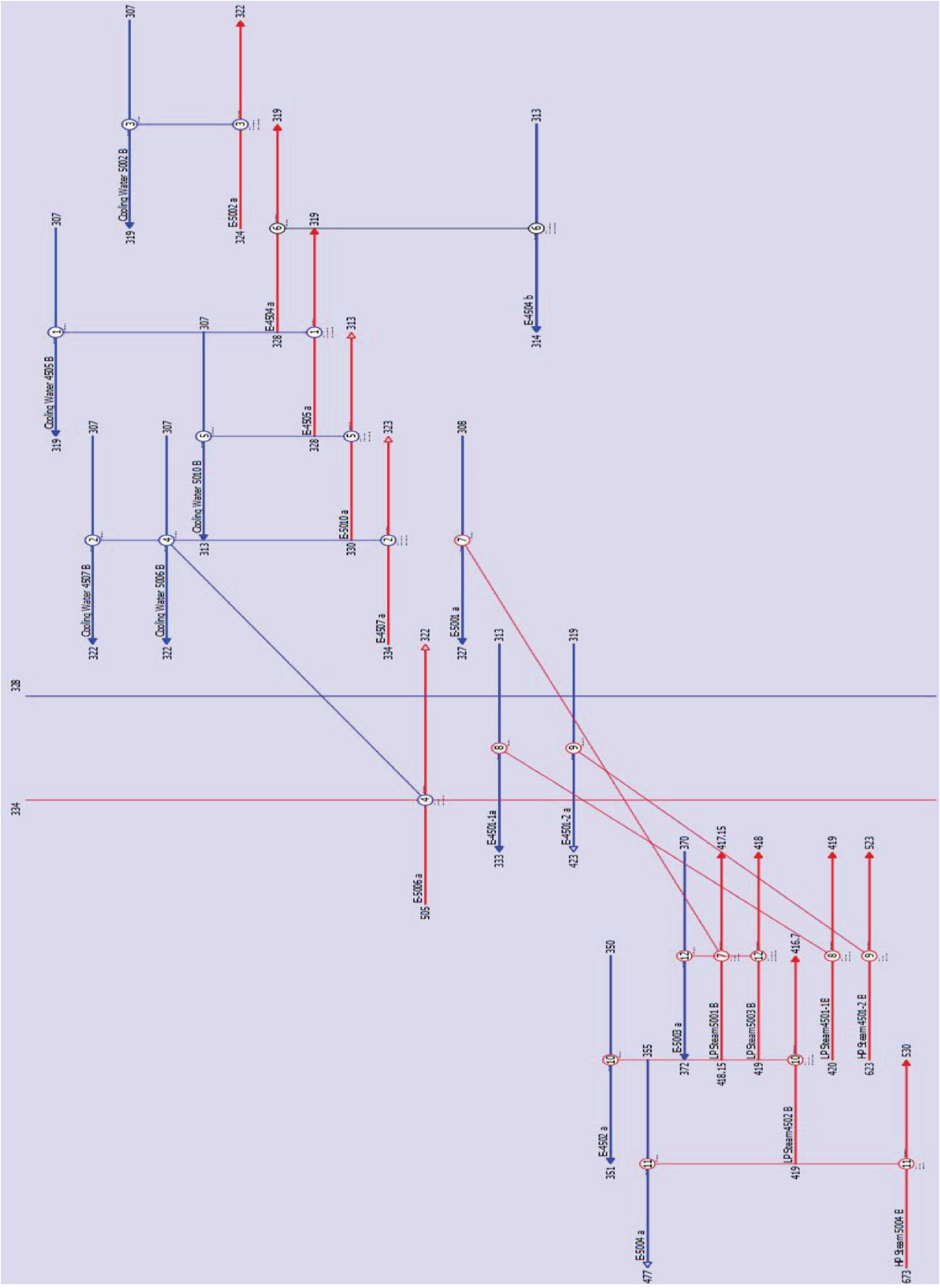

Figure 5.5: A-1 Pinch configuration for existing network

Indeed, when pinch analysis was applied to the current heat exchangers network, 4 heat exchangers out of the 12 (E-4501-2, E-4501, E-5001, E-5006) crossed the pinch temperature as shown in Table (5.5) and figure (5.A-1 support document).

Table 5.5: Heat exchangers that across the pinch and cross pinch duty for existing network

No	Name	Duty [kW]	Cross pinch duty[kW]	Hot side (Pinch Temperature:356)			
				Stream Name	MCp [kW/K]	Inlet Temp [K]	Outlet Temp [K]
9	E-4501-2	112.014	9.52	HP Steam 4501-2 B	1.13	623	523
8	E-4501	2551.95	1891.9	LP Steam 4501-1B	2551.95	420	419
7	E-5001	598.52	598.52	LP Steam 5001 B	598.52	418.15	417.15
4	E-5006	1376.58	1197.33	E-5006 a	7.002	505	308.4

5.5 Revamping Optimization

After identifying the bottleneck from the pinch configuration, we attempt to revamp the networks using the available optimization techniques. The optimum choice is obtained by adding new additional area, new matches, re-allocating existing matches within the various constrains such as maximum added area, space limits. We will search for a cost effective as well as a practical revamping solution.

The search of this optimum is based on two methods:

5. fixed structured and
6. Simulated Annealing.

Figure (5.5) shows the detailed block diagram of the retrofitting procedure using i-heat commercial software.

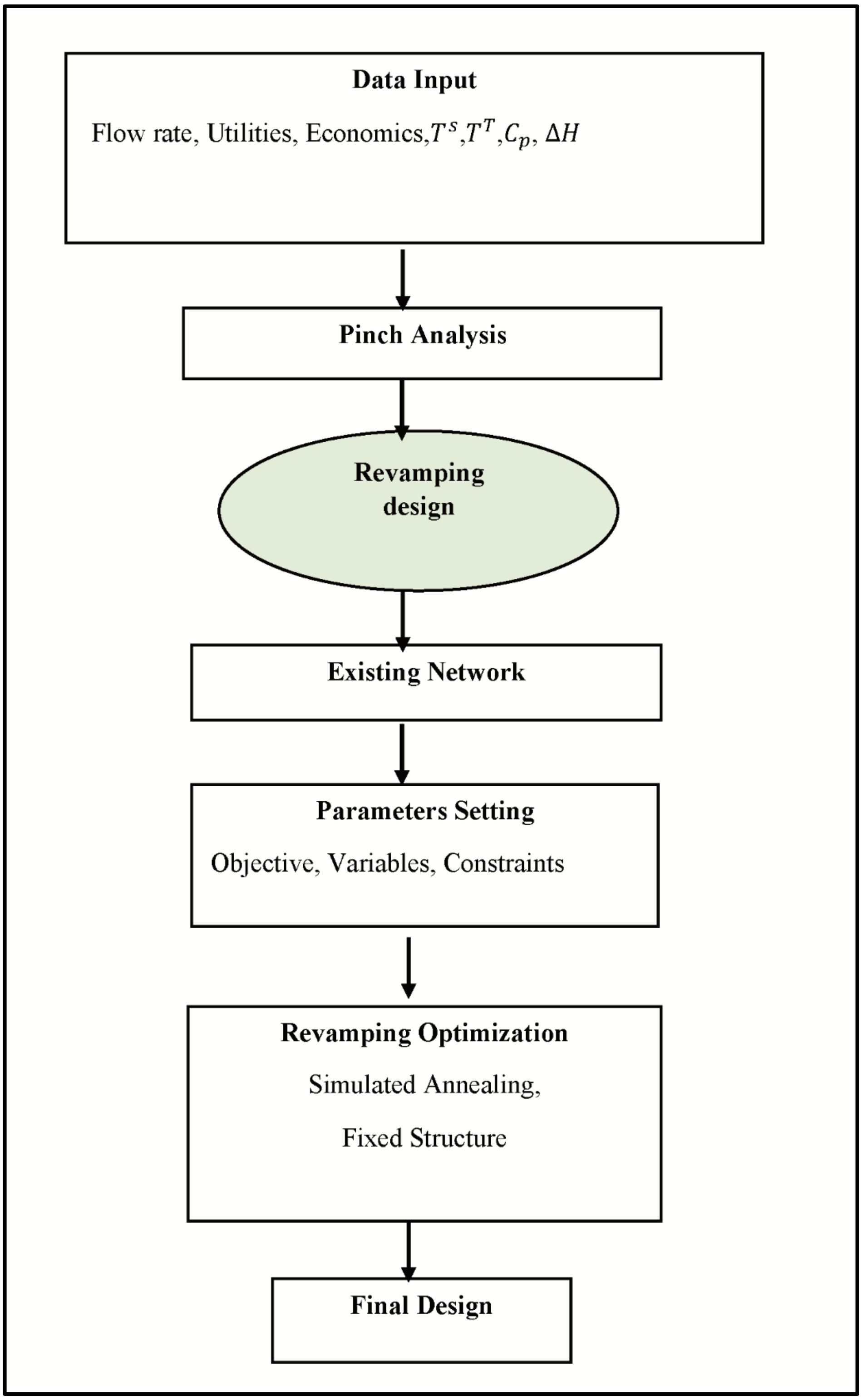

Figure 5.6: The block diagram showing Retrofitting Procedure

5.6 Design options:

5.6.1 Simulated Annealing [SA]

Stochastic optimization in which the structure being optimized is randomly moved from one state to another state by series of defined moves. The moves available depend on the nature of structure being optimized. We have many steps to follow, including:

a. **General**: In this step, we select our objective function, minimizing (TAC) as follows:

   The objective: *Min Total Annual Cost.*

   Total Annualized Cost (TAC) = Annualized Energy cost + Annualized Capital Cost

b. **Enthalpy change (ΔH):** Used to table into account the allowed process stream variations in revamping optimization. The function can be used to analyze the energy effect of background process change. We selected Stream ΔH on the panel. Then include the stream in optimization and specify the lower and upper bounds of enthalpy change ΔH.

c. **Constraints and global network constraints**: In this step, the following constraints can be specified:

   - Limited number of new heat exchanger
   - Limited number of re-sequencing
   - Limited number of re-piping
   - Limited number of splitter

In order to carry out the optimization of the HEN, we did some modifications according to the constraints by the pinch configuration. Five heat exchangers were added namely 15, 17, 18, 19, and 20. Thus, the total number of heat exchangers in the retrofit design become 17. Also, the stream of HE-4504b was splitting to two streams. Finally, one heat exchanger was re-sequenced namely [6]. and ten heat exchangers were re-piped namely [1,2,3,4,5,7,8,9,10,11]. The details of the re-piped streams are listed in Table (5.6). The comparison of the existing and retrofit HEN is shown schematically in Figure (5.6).

Table 5.6: The re-piped heat exchanger Network

Name of re-piped HEs	Existing (before revamping) the connection between:	After revamping the connection between:
HE 1	Stream E-4505a and stream E-4505 B (cooling water)	Stream E-4505a and stream E-4501-1a
HE 2	Stream E-4507 B (cooling water) and Stream E-4507a	Stream E-5002a and stream E-5001a
HE 3	Stream E-5002B (cooling water) and Stream E-5002a	Stream E-4505a and stream E-4504B
HE 4	Stream E-5006a and stream E-5006B (cooling water)	Stream E-5006a and stream E-5003a
HE 5	Stream E-5010B (cooling water) and stream E-5010a	Stream E-5002a and stream E-4501-1a
HE 7	Stream E-5001a and stream E-5001B (LP steam)	Stream E-4502a and stream E-5001B (LP steam)
HE 8	Stream E-4501-1a Stream E-4501-1B (LP steam	Stream E-4507a and stream E-4504B (Splitting)
HE 9	Stream E-4501-2a and stream E-4501-2B (HP steam)	Stream E-5002a and stream E-4504B (Splitting)
HE 10	Stream E-4502a and stream E-4502B (HP steam)	Stream E-5002a and stream E-4504B
HE 11	Stream E-5004a and stream E-5004B (HP steam)	Stream E-5004a and stream E-5006a

Table 5.7: Heat exchangers that across the pinch and cross pinch duty for Simulated Annealing network

Stream	Previous Exchanger					Next Exchanger			
	Hot Stream	MCp[kW/K]	Exchanger	Serial Number	Inlet Temp[K]	Hot Stream	Exchanger	Serial Number	Inlet Temp[K]
E-5003 a	E-5006 a	7.01	E-5006	4	482.14	LP Steam 5003 B	E-5003	12	418.95
Cooling Water 5006 B	E-5010 a	58.45	EX17516	19	330	E-4504 a	EX16842	20	327.86

Table 5.8: Heat exchangers that Criss Cross heat for Simulated Annealing

No	Name	Duty [kW]	Cross pinch duty[kW]	Hot side (Pinch Temperature:356)			
				Stream Name	MCp [kW/K]	Inlet Temp [K]	Outlet Temp [K]
18	EX15239	378.11	294.08	E-5006 a	7.0	376	322
17	EX16816	1387.39	756.76	LP Steam 5003 B	29849.6	419	418.95
15	EX16602	110.04	9.52	E-5006 a	7.0	497.85	482.14

Indeed, when Simulated Annealing was applied, to the current heat exchangers network, the duty of HEs that crossed the pinch as shown in figure (5. A-2 support document), was reduced as shown in Table (5.7). Moreover, the criss-cross heat transfer has been found for the hot stream heat exchanger as shown in Table (5.8).

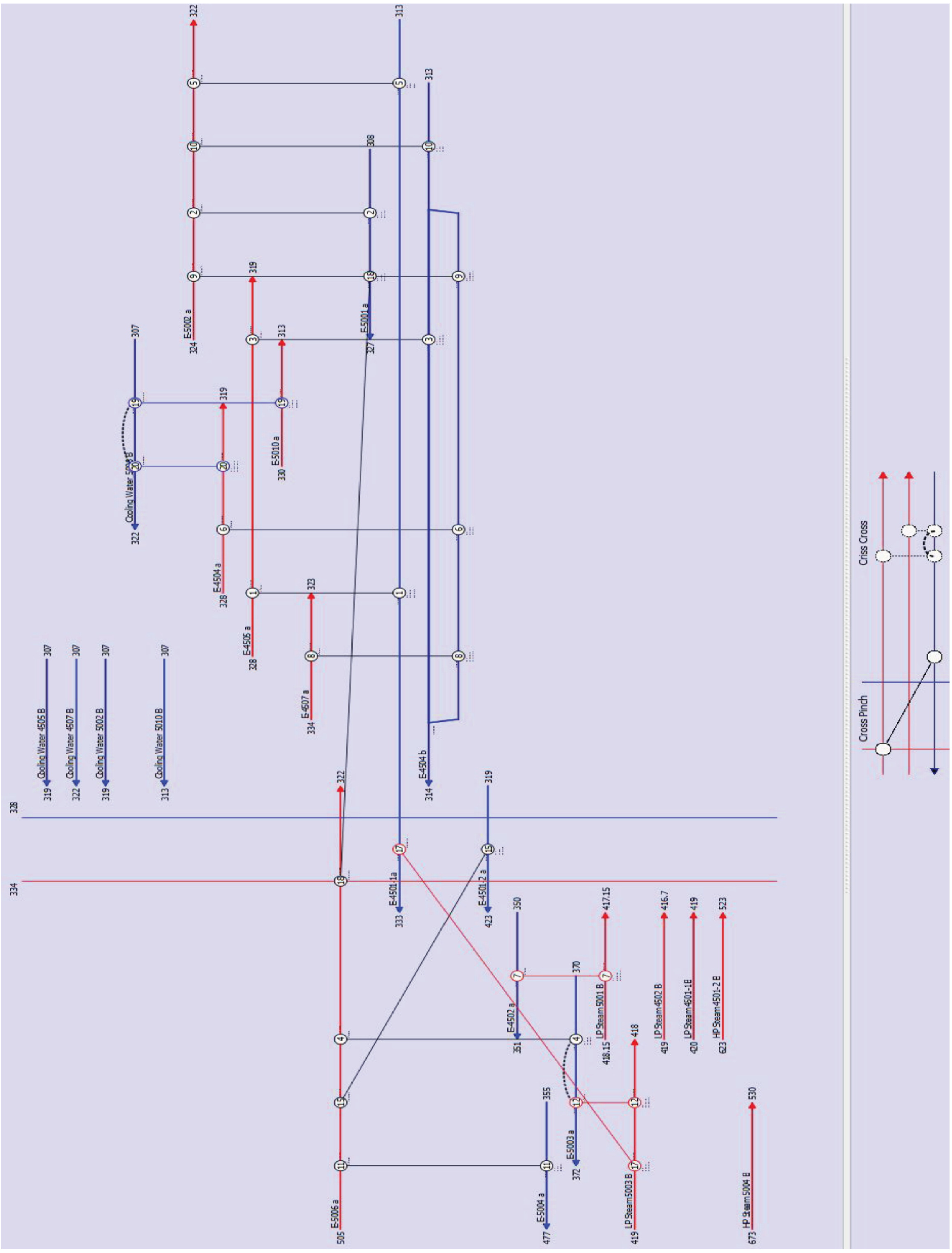

Figure 5.7: A-2 Pinch configuration for Simulated Annealing (SA)

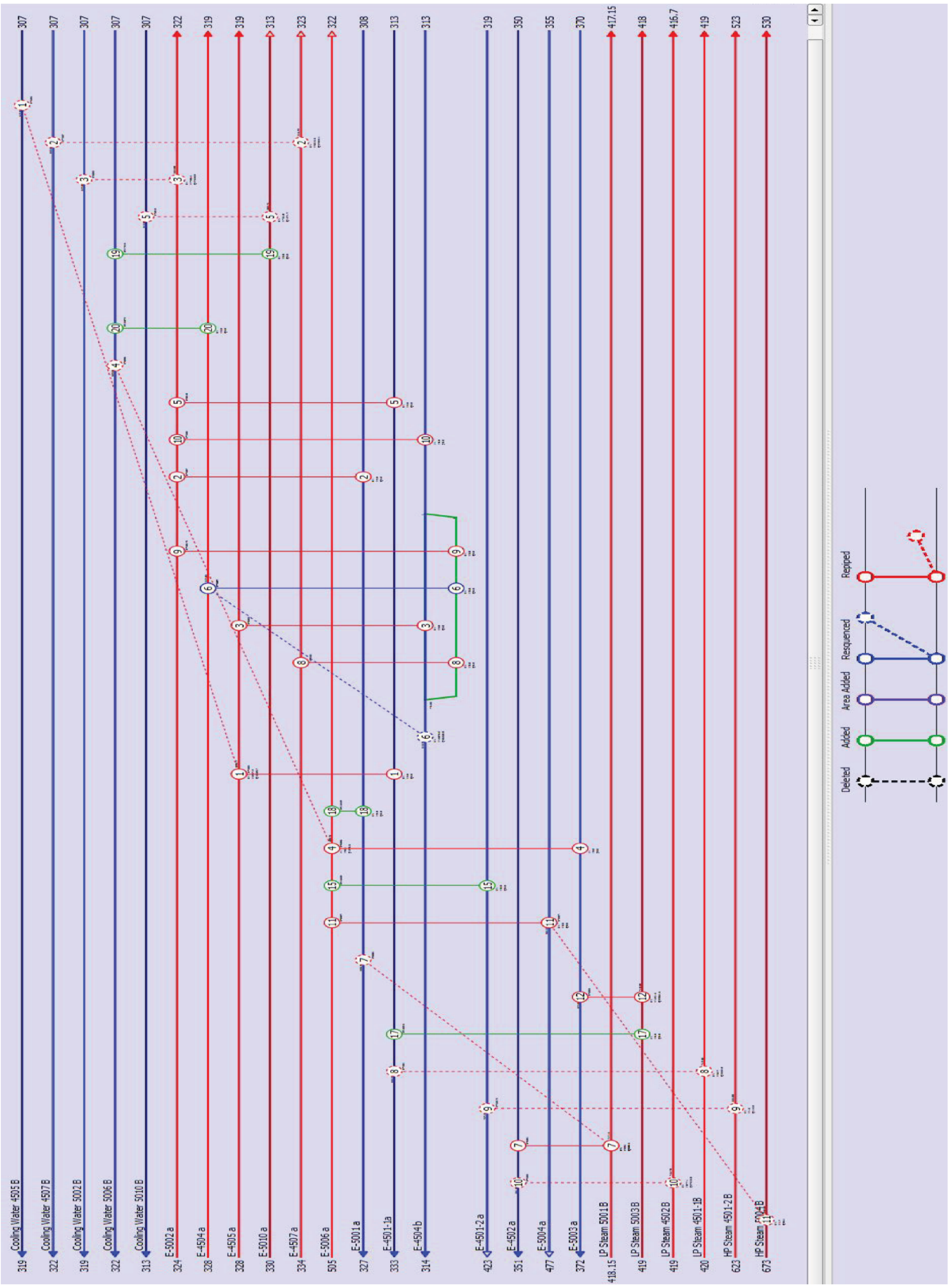

Figure 5.8: Superstructure showing a comparison between the existing network and the retrofitted one by Simulated Annealing (SA)

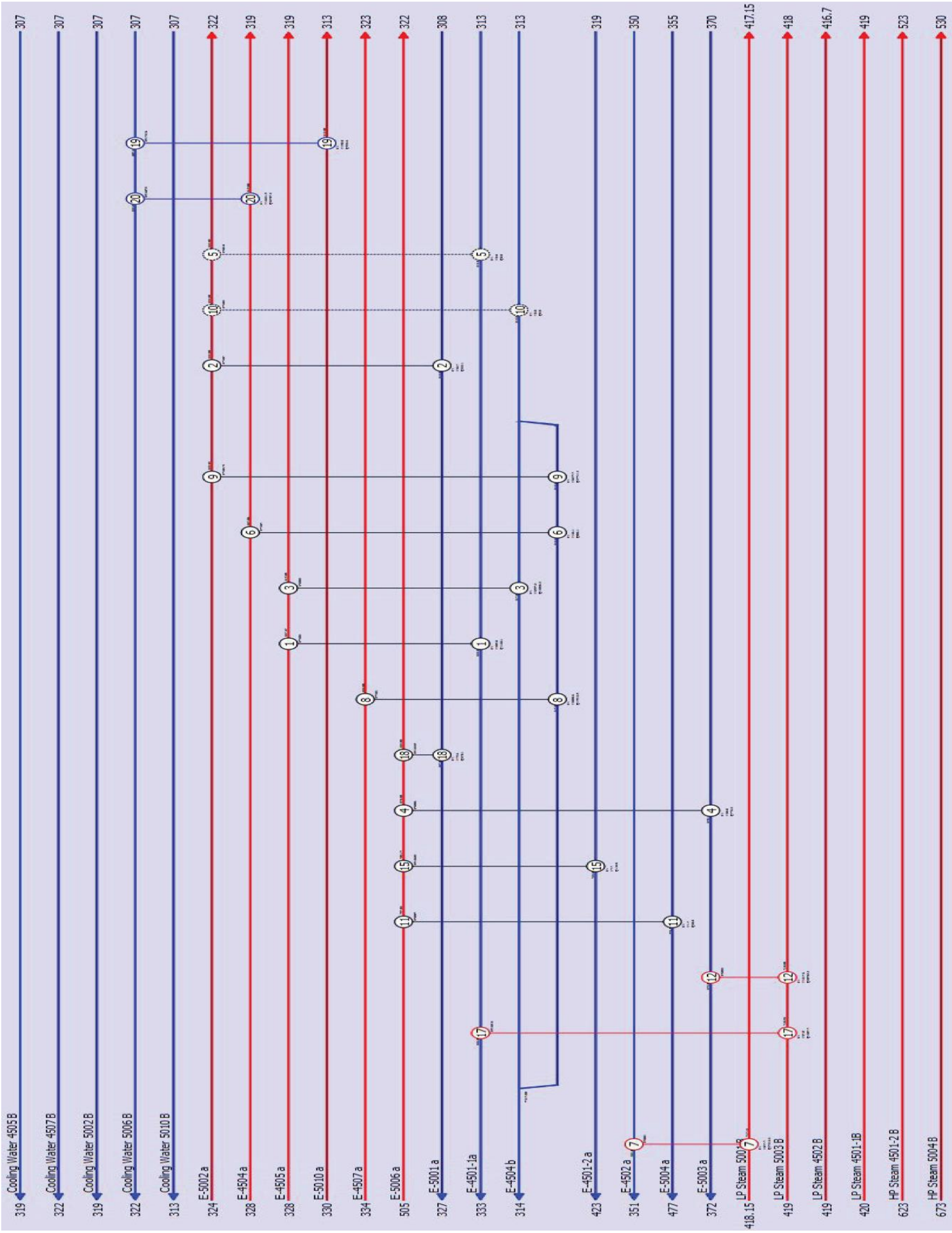

Figure 5.9: The final design after optimization using Simulated Annealing [SA]

Table 5.9: Summary of result of Simulated Annealing [SA] (comparison between the Existing Network and Retrofitting network)

Items		Retrofitting Network	Existing Network
ΔT_{min}	K	6.00	6.05
Recover ratio		52.95	46.70
Hot utility duty	kW	54062.91	57419.19
Cold utility duty	kW	56780.69	74675.46
Number of exchangers		17	12
Number of shells		17	12
Used area	m^2	32376.79	33676.75
Total Annual cost	MM$/y	23.3891	26.0056
Total capital cost	MM$/y	4.0375	5.1803
Heat exchanger capital cost	MM$/y	0.6343	0.6509
Total operating cost	MM$/y	19.3516	20.8253
Hot utility operating cost	MM$/y	10.5161	11.1786
Cold utility operating cost	MM$/y	5.6102	8.6551
Other operating cost	MM$/y	3.2253	0.9917

The results of the retrofit design by simulated annealing are summarized in Table (5.9) and shown in figure (5.7). Here, we here found that the Total Annual Cost is reduced from 26.01 MM$/y to 23.39 MM$/y, which constitutes a saving of up to 10.07 %.

5.6 2 Fixed structure:

The second optimization technique, that we have employed, to minimize the annual operating cost, is called fixed structure method. In this method, the number of the heat exchanger was not changed, as SA, this leads some modification in topology to get saving in utilities by varying: (1) hot and cold stream rates, (2) heat exchangers duties and areas.

We found that 4 heat exchangers cross the pinch as shown in figured (5.A-3 support document) [E-4501-2, E-4501, E-5001 and E-5006] as shown in Table (5.10).

Indeed, when Fixed Structured was applied, to the current heat exchangers network, the duty of HE, that crossed the pinch, was reduced.

The modified superstructure is shown in figure (5.8). This superstructure, obtained with the fixed structure method, reduces the annual operating cost. The final design for the fixed structure shown in figure (5.9) and table (5.11) summarizes the reduction of Total Annual Cost using this technique. In effect, the TAC is reduced from 26.01 MM$/y to 20.43 MM$/y (i.e., up to 20.01 % saving is achieved).

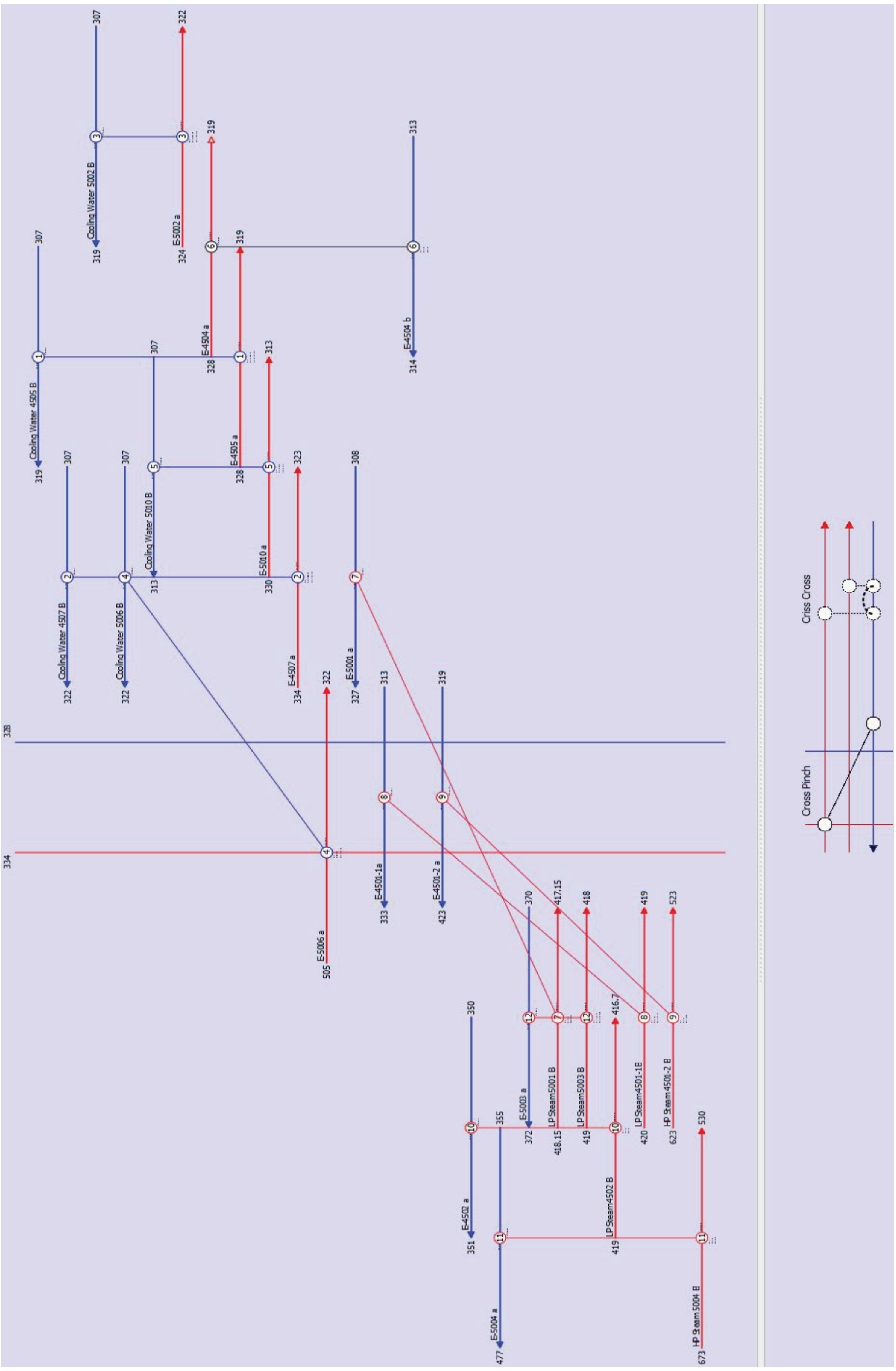

Figure 5.10: A-3 Pinch configuration for fixed structure

Table 5.10: Heat exchangers that across the pinch and cross pinch duty for fixed structure network

No	Name	Duty [kW]	Cross pinch duty[kW]	Hot side (Pinch Temperature:356)			
				Stream Name	MCp [kW/K]	Inlet Temp [K]	Outlet Temp [K]
9	E-4501-2	110	9.52	HP Steam 4501-2 B	1.1	623	523
8	E-4501	2522.54	1891.9	LP Steam 4501-1B	2522.54	420	419
7	E-5001	600.19	600.19	LP Steam 5001 B	600.19	418.15	417.15
4	E-5006	1281.36	1197.33	E-5006 a	7.002	505	322

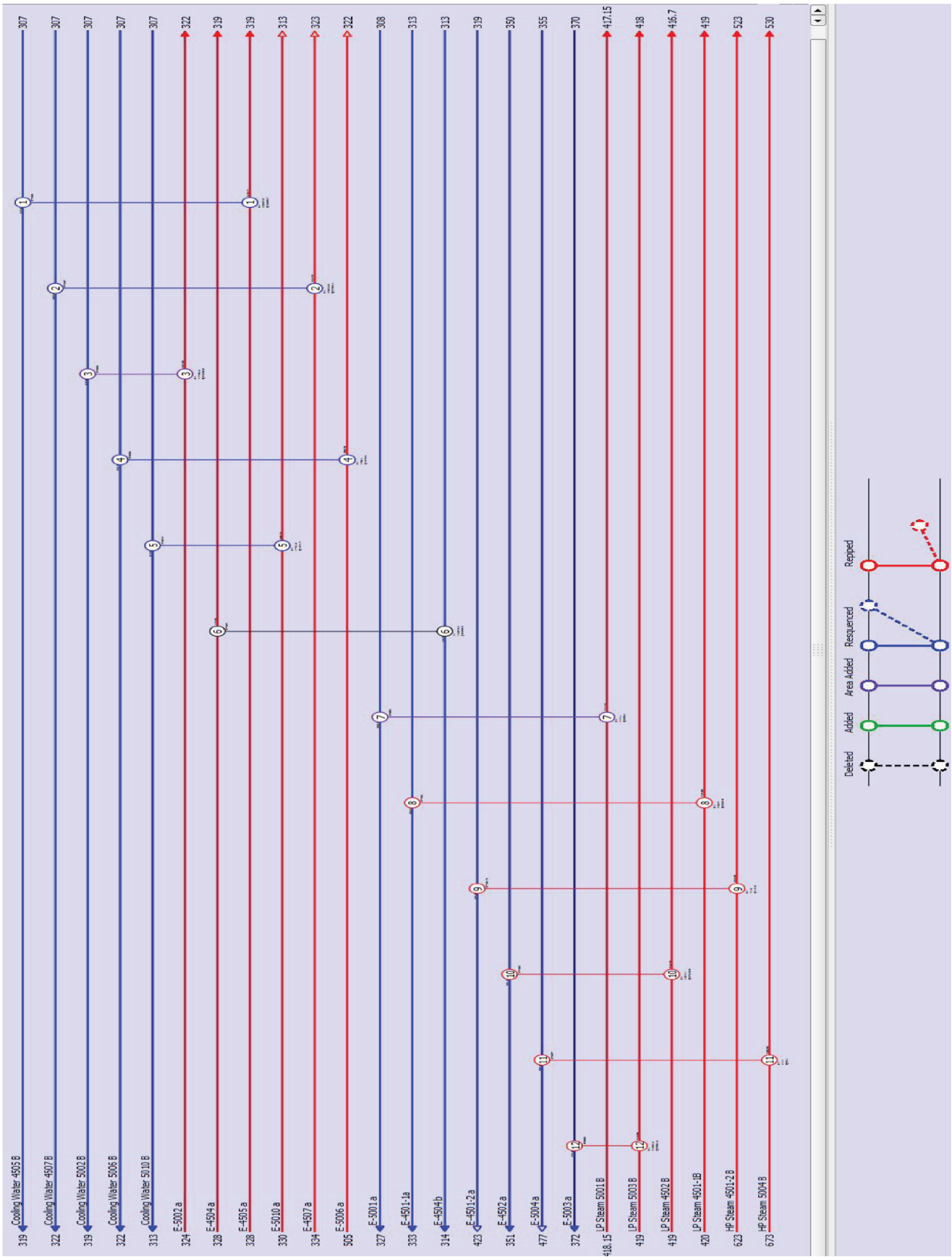

Figure 5.11: Superstructure show the comparing between the existing and fixed structure design

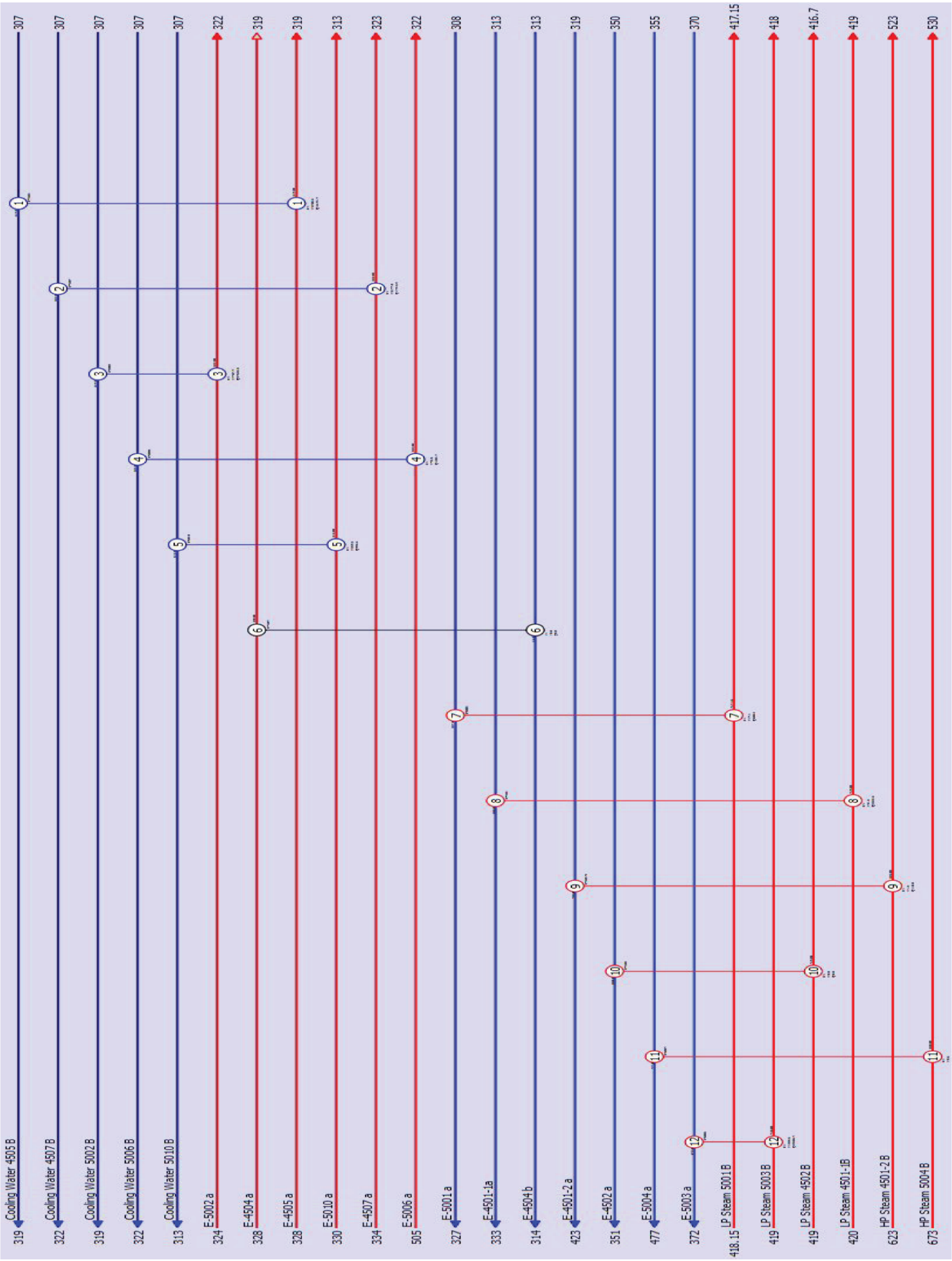

Figure 5.12: Final design for fixed structure

Table 5.11: Summary of result for fixed structure (comparison between the Existing and retrofitting network)

Items		Retrofitting network	Existing Network
ΔT_{min}	K	15.00	6.05
Hot utility duty	kW	32488.18	57419.19
Cold utility duty	kW	64015.40	74675.46
Number of exchangers		12	12
Used area	m^2	16249.26	33676.75
Total Annual cost	MM$/y	20.4270	26.0056
Total capital cost	MM$/y	3.7684	5.1803
Heat exchanger capital cost	MM$/y	0.3530	0.6509
Total operating cost	MM$/y	16.6586	20.8253
Hot utility operating cost	MM$/y	6.3273	11.1786
Cold utility operating cost	MM$/y	7.5549	8.6551
Other operating cost	MM$/y	2.7764	0.9917

5.7 Conclusions

A retrofit design methodology is proposed, for Fractionation unit using the pinch analysis technique, is energy efficient and can be used to find the optimum HEN. A reduction of the total Annual cost (TAC) was obtained by optimizing the HEN for the unit. A revamp design was suggested based on the Simulated Annealing (SA) and the fixed structure techniques. The results of SA optimization were better than the fixed structure. The suggested modern design has a saving of 5.58 million US $/year (save up to 21.45%), over the current design.

NOMENCLATURE

Symbols	Meaning (Units)
A	Area of heat exchanger
E- a & E- B	Denote the process stream and utility stream in HENs
CP	Heat capacity flow rate (MW/K)
Cp	Specific heat capacity (Kj/Kg.K)
F*Cp	Heat capacity flow rate (MW/K)
ΔH	Exchanger duty (MW)
HEN	Heat exchanger network
HENs	Heat exchanger networks
HEs	Heat exchangers
MM	Millions
Q	Duty
R	Interest rate (%)
T^T	Outlet temperature of stream (K)
T^S	Inlet temperature of stream (K)
U	Overall heat transfer coefficient (MW/m2 K)

Appendix A

Summary for stream after Simulation Annealing [SA]

No	Name	Mass Flow	T^S	T^T	ΔH
		kg/h	K	K	kW
2	E-4504 a	660913	328	319	56673.29
4	E-4505 a	189761	328	319	16191.36
5	E-4507 a	303802	334	323	17915.88
7	E-5002 a	343793	324	322	27633.32
10	E-5006 a	6414	505	322	1281.36
11	E-5010 a	64142	330	313	993.49
1	E-4501-1a	166321	313	333	2522.54
3	E-4504 b	2207244	313	314	61269.41
6	E-5001 a	6414	308	327	600.19
8	E-5003 a	1954991	370	372	29205.39
9	E-5004 a	201	355	477	50.07
12	E-4502 a	614811	350	351	24213.31
13	E-4501-2 a	250	319	423	110.00

Appendix B

Utility after Simulation Annealing [SA]

No	Name	T^S	T^T	Duty	Operation Cost Index	Annual Operation Cost	Capital Cost
		K	K	kW	MM$/Kj	MM$/y	MM$
1	LP Steam 4501-1B	420	419	0.00	6.283E-06	0.0000	0.0334
4	LP Steam 5001 B	418	417	24213.31	6.283E-06	4.7099	3.2642
6	LP Steam 5003 B	419	418	29849.60	6.283E-06	5.8062	3.8530
7	HP Steam 5004 B	673	530	0.00	7.860E-06	0.0000	0.0334
10	LP Steam 4502 B	419	417	0.00	6.283E-06	0.0000	0.0334
11	HP Steam 4501-2 B	623	523	0.00	7.860E-06	0.0000	0.0334
2	Cooling Water 4505 B	307	319	0.00	3.989E-06	0.0000	0.0334
3	Cooling Water 4507 B	307	322	0.00	3.191E-06	0.0000	0.0334
5	Cooling Water 5002 B	307	319	0.00	3.989E-06	0.0000	0.0334
8	Cooling Water 5006 B	307	322	56780.69	3.191E-06	5.6102	6.4222
9	Cooling Water 5010 B	307	313	0.00	7.979E-06	0.0000	0.0334

Appendix(C)

Stream for fixed structure after retrofitting

No	Name	Mass Flow	T^S	T^T	ΔH
		kg/h	K	K	kW
2	E-4504 a	660913	328	319	56673.29
4	E-4505 a	189761	328	319	16191.36
5	E-4507 a	303802	334	323	17915.88
7	E-5002 a	343793	324	322	27633.32
10	E-5006 a	6414	505	322	1281.36
11	E-5010 a	64142	330	313	993.49
1	E-4501-1a	166321	313	333	2522.54
3	E-4504 b	2207244	313	314	61269.41
6	E-5001 a	6414	308	327	600.19
8	E-5003 a	1954991	370	372	29205.39
9	E-5004 a	201	355	477	50.07
12	E-4502 a	614811	350	351	24213.31
13	E-4501-2 a	250	319	423	110.00

Appendix (D)

Utility after fixed structure

No	Name	T^S	T^T	Duty	Operation Cost Index	Annual Operation Cost	Capital Cost
		K	K	kW	$/kJ	MM$/y	MM$
1	LP Steam 4501-1B	420	419	2522.54	6.283E-06	0.4907	0.5625
4	LP Steam 5001 B	418	417	600.19	6.283E-06	0.1167	0.2012
6	LP Steam 5003 B	419	418	29205.39	6.283E-06	5.6809	3.7869
7	HP Steam 5004 B	673	530	50.07	7.860E-06	0.0122	0.0564
10	LP Steam 4502 B	419	417	0.00	6.283E-06	0.0000	0.0334
11	HP Steam 4501-2 B	623	523	110.00	7.860E-06	0.0268	0.0766
2	Cooling Water 4505 B	307	319	16191.36	3.989E-06	1.9998	2.3749
3	Cooling Water 4507 B	307	322	17915.88	3.191E-06	1.7702	2.5724
5	Cooling Water 5002 B	307	319	27633.32	3.989E-06	3.4130	3.6243
8	Cooling Water 5006 B	307	322	1281.36	3.191E-06	0.1266	0.3412
9	Cooling Water 5010 B	307	313	993.49	7.979E-06	0.2454	0.2845

Appendix E

Matching after Simulation Annealing [SA]

No	Tag	Stream	Mass Flow	Inlet T	Outlet T	LMTD	Q	A	Added A	Capital Cost
			kg/h	K	K	K	kW	m²	m²	MM$
1	E-4505	E-4505 a	189761.00	328	327	9.58	1135.14	289.0	289.0	0.0000
		E-4501-1a	166321.00	313	322					
2	E-4507	E-5002 a	343793.00	322	322	10.09	222.08	53.7	53.7	0.0000
		E-5001 a	6414.00	308	315					
3	E-5002	E-4505 a	189761.00	327	319	9.20	15056.22	3994.2	3994.2	1.4610
		E-4504 b	542403.48	313	314					
4	E-5006	E-5006 a	6414.00	482	376	36.24	743.18	50.0	50.0	0.3390
		E-5003 a	1954991.00	370	370					
5	E-5010	E-5002 a	343793.00	322	322	9.00	0.00	0.0	0.0	0.0014
		E-4501-1a	166321.00	313	313					

6	E-4504	E-4504 a	660913.00	328	328	14.33	886.09	136.1	136.1	0.0246
		E-4504 b	1664840.52	314	314					
8	E-4501	E-4507 a	303802.00	334	323	14.03	17915.88	3000.6	3000.6	0.0344
		E-4504 b	1664840.52	314	314					
9	E-4501-2	E-5002 a	343793.00	324	322	9.69	27411.23	6644.4	6644.4	0.0001
		E-4504 b	1664840.52	313	314					
10	E-4502	E-5002 a	343793.00	322	322	9.00	0.00	0.0	0.0	0.4959
		E-4504 b	2207244.00	313	313					
11	E-5004	E-5006 a	6414.00	505	498	70.55	50.03	1.7	1.7	0.4263
		E-5004 a	201.00	355	477					
15	EX16602	E-5006 a	6414.00	498	482	113.30	110.04	4.4	4.4	0.8505
		E-4501-2 a	250.00	319	423					
18	EX15239	E-5006 a	6414.00	376	322	21.55	378.11	79.3	79.3	0.0155
		E-5001 a	6414.00	315	327					
7	E-5001	LP Steam 5001 B	37422384.84	418	417	67.15	24213.31	847.4	847.4	0.0246
		E-4502 a	614811.00	350	351					

12	E-5003	LP Steam 5003 B	48761500.97	419	418	47.45	28462.21	1517.6	1517.6	0.0250
		E-5003 a	1954991.00	370	372					
17	EX16816	LP Steam 5003 B	48761500.97	419	419	91.37	1387.39	37.8	37.8	0.1112
		E-4501-1a	166321.00	322	333					
19	EX17516	E-5010 a	64142.00	330	313	12.56	993.49	198.8	198.8	0.1175
		Cooling Water 5006 B	3261755.79	307	307					
20	EX16842	E-4504 a	660913.00	328	319	8.46	55787.20	15521.9	15521.9	0.1039
		Cooling Water 5006 B	3261755.79	307	322					

Appendix F

Matching after Fixed Structure

No	Tag	Stream	Mass Flow	Inlet T	Outlet T	LMTD	Q	A	Added A	Capital Cost
			kg/h	K	K	K	kW	m^2	m^2	MM$
6	E-4504	E-4504 a	660913.00	328	328	15.00	0.01	0.0	0.0	0.0100
		E-4504 b	2207244.00	313	313					
1	E-4505	E-4505 a	189761.00	328	319	10.43	16191.36	3788.5	3788.5	0.5933
		Cooling Water 4505 B	1162660.00	307	319					
2	E-4507	E-4507 a	303802.00	334	323	13.90	17915.88	3144.0	3144.0	0.5124
		Cooling Water 4507 B	1029174.19	307	322					
3	E-5002	E-5002 a	343793.00	324	322	9.10	27633.32	7407.4	7407.4	1.0073
		Cooling Water 5002 B	1984277.91	307	319					

4	E-5006	E-5006 a	6414.00	505	322	67.16	1281.36	46.6	46.6	0.0273
		Cooling Water 5006 B	73607.30	307	322					
5	E-5010	E-5010 a	64142.00	330	313	10.56	993.49	229.5	229.5	0.0719
		Cooling Water 5010 B	142678.24	307	313					
7	E-5001	LP Steam 5001 B	927611.56	418	417	99.88	600.19	14.1	14.1	0.0167
		E-5001 a	6414.00	308	327					
8	E-4501	LP Steam 4501-1B	3933712.35	420	419	96.19	2522.54	61.6	61.6	0.0316
		E-4501-1a	166321.00	313	333					
9	E-4501-2	HP Steam 4501-2 B	1420.68	623	523	201.99	110.00	1.3	1.3	0.0110
		E-4501-2 a	250.00	319	423					
10	E-4502	LP Steam 4502 B	21.43	419	419	69.00	0.00	0.0	0.0	0.0100
		E-4502 a	614811.00	350	350					

11	E-5004	HP Steam 5004 B	482.70	673	530	185.30	50.07	0.6	0.6	0.0106
		E-5004 a	201.00	355	477					
12	E-5003	LP Steam 5003 B	47709139.34	419	418	47.50	29205.39	1555.6	1555.6	0.2962
		E-5003 a	1954991.00	370	372					

CHAPTER 6

Simulating and Applying Aspen Energy Analyzer for Energy Savings Enhancement

6.1 Introduction

Nowadays in the process industry, the optimization and design procedures are more inclined towards the minimization of energy consumption. Energy saving has both economic and environmental benefits [1]. The use of Pinch Technology can help in finding a balance between energy and economics as well as defining the correct location of utilities and heat exchangers. The procedure first predicts the minimum requirements of external energy network area and a number of units for given process at the pinch point. In the second stage, the heat exchanger network design that satisfies these targets is proposed [44][45]. The energy optimization techniques can be used to determine the thermodynamic properties [2]. The occurrence of energy crises has drawn the attention people of industry and academia to start striving for an effective research and development in the fields of economic utilization of energy in process industries[3]. This situation, coupled with growing concern for the pollution-free environment, has resulted in research efforts in energy optimization and best process integration for optimum use of energy a welcome notion all over the world[4]. The tools such as pinch analysis have been applied widely to analyze the process plants, including petrochemicals. However, some works on pinch analysis strived to generate optimal heat exchanger network for the systems and development of mathematical models to minimize Total Annual Cost (TAC) with some of the assumptions made were relied on constant process parameters [5]–[8]. The assumptions of constant process streams properties result in conclusions that are far away from industrial realities. However, there is increasing attention toward the application of pinch concepts in the models to reduce the total annual operating cost [9].

Generally, the purpose of process optimization is to increase the plant throughput by operating the plant at or near to the optimum conditions. Combining all these concepts, along with their usefulness to optimal and cost effective operation of petrochemical, this work strives to bridge the existing gaps in the previous studies by conducting analysis of petrochemical units, establishing generic optimum conditions based on varying process properties rather than the assumption of constant scenarios, and quantifying the extent of process streams' inefficiencies that were cost plant operator and concept. High competition in the ever dynamic international market and stringent product requirements are driving force compelling the process industries to look keenly into plant operation in a profitable manner. To avert the possible dwindling in profit margin due to high competition and to maintain the share in the international market arena, the need for real-time or on-line optimization of the entire plant is equally essential and hence covered in this study.

According to the current report by the Organization of the Petroleum Exporting Countries, Saudi Arabia possessed 18 percent of the world's proven petroleum reserves and ranked as the largest exporter of petroleum. For this country to maintain its exalted position in OPEC, the need of conducting research that can improve the performance of the existing facilities is justified and should be in high demand. Globally, the general energy supply and environmental situation require an improved management for the utilization of energy resources. Improved petrochemical plant efficiency for optimum production at minimum energy consumption can reduce carbon dioxide emissions associated with fossil-fuel combustion. In the global sense, this research work also has the environmental protection aspect which is a hot topic of discussion worldwide today.

The specific case study does not intend to start the design of petrochemical industry from the scratch; rather, it focuses on the analysis of the functional petrochemical operations in the Kingdom of Saudi Arabia. The process data for the analysis was obtained from the petrochemical in the Kingdom through exploration of the existing linkages between KFUPM and the petrochemical industries in the Kingdom. This work further aims to specify some basic thermodynamic parameters for optimum performance of the plant that was missing in the previous literature and the detailed, energy, and economic analysis of the plant.

In this study, heat integration in Aspen Energy Analyzer was considered for analyzing and developing the performance of Heat Exchanger Network (HEN). HEN operations features are designed to provide an understanding of current plant operation. HEN design features assist the designer in understanding the gap between current operation and the thermodynamic optimum operation. Furthermore, Aspen Energy Analyzer was used to identify and compare options to improve the performance and reduce the gap between current and thermodynamic optimum operations.

Finally, heat transfer intensification and network utility exchanger modification algorithm are considered simultaneously to achieve more energy saving in the existing HENs and whilst reduce retrofit cost significantly. The findings of this study will help in effective energy saving of the entire plant, processed-bottlenecking, investment cost reduction, process modification and total site planning.

In this work, we optimize the design after network evaluation using the Aspen Analyzer software and the pinch technique by modifying the utility heat exchanger method to trade off capital cost with energy recovery.

6.2 Process Description and Existing heat exchanger network

Stream data extracted, which in turn dictate the HEN. The current process consists of 12 heat exchangers. There is a total of 24 streams. 7 cold process streams and 6 hot process streams as well as 6 hot and 5 cold utility streams. The hot exchanger's utility uses either HP or LP steam the cold utility heat exchangers use is cooling water. The data provided from the plant have been simulated the process, which included the temperature, heat capacity, enthalpy and flow rate of streams. The details of all the streams and utilities are listed in Table (6.1) and (6.2), respectively. The grid diagram of an existing heat exchanger network of fractionation unit in the plant under study is shown in figure (6.1).

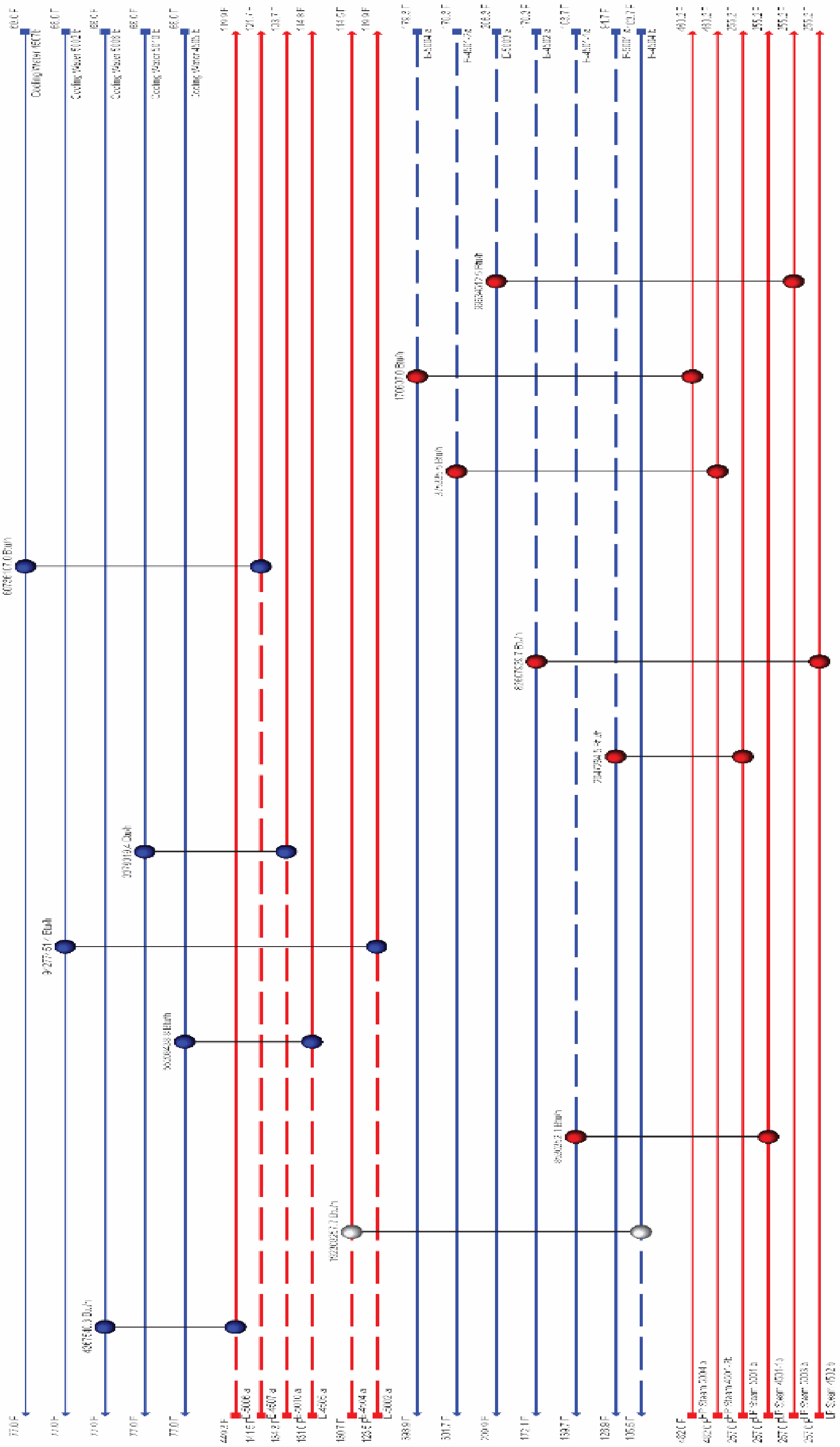

Figure 6.1: Heat exchanger Network for existing design

Table 6.1: The hot and cold stream for existing design

Name	Inlet T [^{o} C]	Outlet T [^{o} C]	Heat Capacity [KJ/^{o} C-h]	Enthalpy [KJ/h-m^2-^{o}C]
E-4501-1a	39.9	59.9	125.00	2369.54
E-4504 a	54.9	45.9	6262.22	53418.94
E-4504 b	39.9	40.9	56360.00	53418.96
E-4505 a	55.0	46.0	1797.78	15335.70
E-4507 a	60.9	49.9	1618.18	16871.12
E-5001 a	34.9	53.9	31.58	568.71
E-5002 a	50.9	48.9	13815.00	26188.18
E-5003 a	96.9	98.9	14600.00	27676.25
E-5004 a	81.9	203.9	0.41	47.41
E-5006 a	231.9	48.9	6.99	1212.42
E-5010 a	56.9	39.9	56.24	906.19
E-4502 a	76.9	77.9	24210.00	22946.65
E-4501-2a	76.9	149.9	1.06	73.34

Table 6.2: The hot and cold utilities for existing design

	Utility	Inlet T [^oC]	Outlet T [^oC]	Effective Cp [KJ/KG-^oC]
1	LP Steam 4501-1b	125.0	124.0	2196.4
2	Cooling Water 4505 b	20.0	25.0	4.183
3	Cooling Water 4507b	20.0	25.0	4.183
4	LP Steam 5001 b	125.0	124.0	2196.4
5	Cooling Water 5002 b	20.0	25.0	4.183
6	LP Steam 5003 b	125.0	124.0	2196.4
7	HP Steam 5004 b	250.0	249.0	1703.1
8	Cooling Water 5006 b	20.0	25.0	4.183
9	Cooling Water 5010 b	20.0	25.0	4.183
10	LP Steam 4502 b	125.0	124.0	2196.4
11	HP Steam 4501-2b	250.0	249.0	1703.1

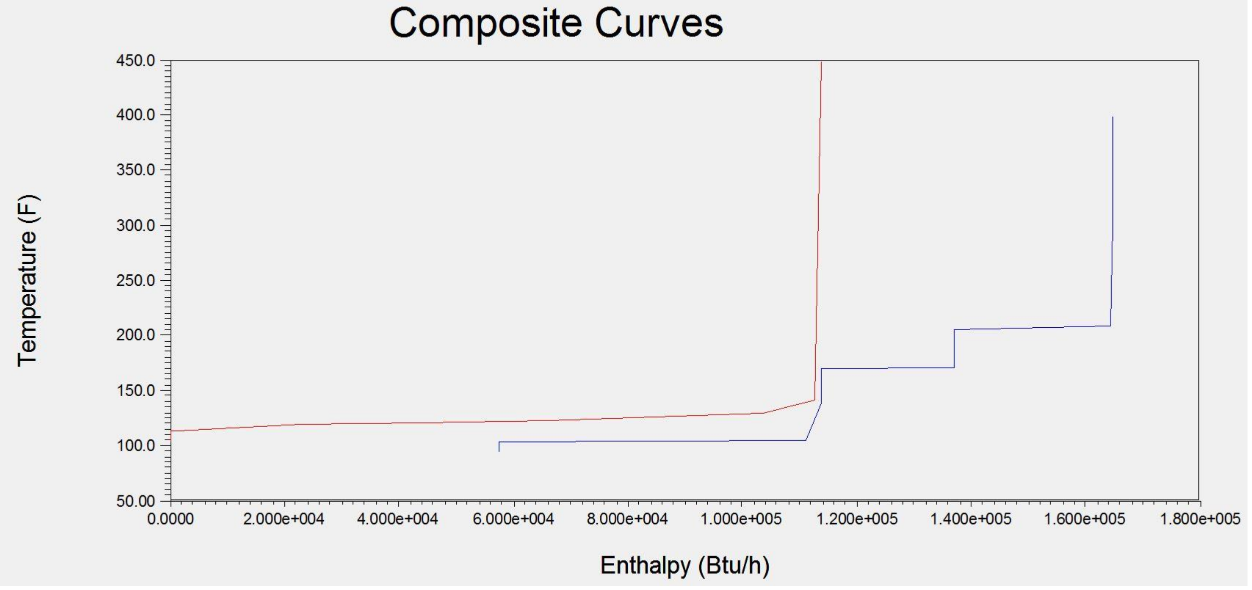

Figure 6.2: Composite Curves for existing Heat Exchanger Network of the Fractionation Unit

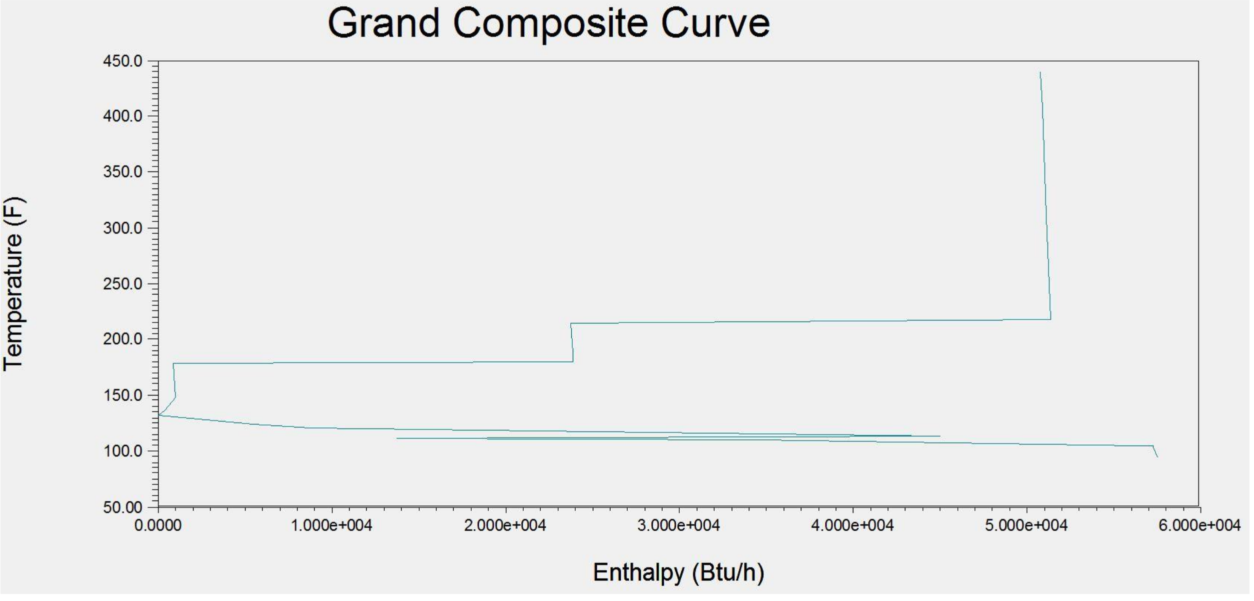

Figure 6.3: The Existing Heat Exchanger Network Grand Composite Curve of the Fractionation Unit

Table 6.3: Information about existing design

Network Cost indexes	
	Cost Index
Heating [MM$/year]	3.40093099
cooling [MM$/year]	0.42828994
Operating [MM$/year]	3.82922094
Capital [MM$]	0.234
Total Cost [MM$/year]	3.90472431
Network Performance	
	HEN
Heating [KW]	56669980.737
cooling [KW]	63879978.299
Number of Unit	12
Number of Shells	52

Table 6.4: Summary for Base case

Heat Exchanger	Area [m^2]	load [KW]
E-103	0.10264	1279.69
E-105	159.582	56266.26
E-110	1.60875	2499.39
E-100	18.7536	16176.07
E-102	111.763	27623.29
E-104	0.66397	989.76
E-111	0.31135	599.85
E-109	151.153	24204.12
E-101	17.0274	17795.68
E-107	9.17E-03	109.97
E-106	5.39E-03	49.99
E-108	117.82	29192.912

Table 6.5: Utility summary

Utility	Cost index [Cost/year]	Load [KW]
LP Steam 4501-1b	149899	2499.39
Cooling Water 4505 b	108480	16176.07
Cooling Water 4507b	119342	17795.68
LP Steam 5001 b	35975.7	599.85
Cooling Water 5002 b	185248	27623.29
LP Steam 5003 b	1750816	29192.91
HP Steam 5004 b	3944.7	49.99
Cooling Water 5006 b	8581.89	1279.69
Cooling Water 5010 b	6637.56	989.76
LP Steam 4502 b	1451618	24204.12
HP Steam 4501-2b	8678.34	109.97

6.3 Cost Data

The correct cost data is essential to take a decision about the economic feasibility of a successful project, therefore an economical evaluation must be performed.

The heat exchanger capital cost index parameters which were measured for the calculation of heat exchanger capital cost index values were as follows: a =100000, b = 800, c= 0.8. The typical operation time is 8765.76 hrs, while the plant is assumed to have a life span of 5 years. The annual interest rate is assigned to be roughly 10%. The mathematical formula, used to determine the annualization factor [38], is given by:

$$AF= (1 + ROR/100)^{(PL / PL)}$$

The process costs are mostly related to the consumption of fuel for heat generation. This heat is necessary for steam generation to supply the hot streams. Furthermore, the cost related to cold utilities is also involved, but it is on the lower side in comparison with that required for the hot utilities

Utility Streams

The Utility Streams lets us to require the utilities used in heat exchanger network to heat or cool the stream in the process. Here we are provided that cooling utility as cooling water is obtainable at 200 ^{o}C and hot utility as a low-pressure steam is offered at 1250 ^{o}C. The cost index of cooling water and the LP steam are and 2.125×10^{-7} and 1.9×10^{-6} respectively.

Capital cost

Capital Cost Index (Cost) = a + b (Area (Heat Exchanger) or Duty (Fired Heater)) ^{C}

The heat exchanger (HE) and utility cost are given by the following equations:

HE cost = $A_1 + B_1$ (area) ^{C_1}

Energy cost (Utility) = $A_2 + B_2$ (Duty) ^{C_2}

Annualized Total Cost (TAC) = Annualized Energy cost + Annualized Capital Cost

Where A represent a fixed cost. B_1, the heat exchanger cost per unit, depend on the type of material.

6.4 Pinch Analysis

Pinch Analysis is used, and it was found that the hot and cold utility demands of the existing network were 56669980.737 [KW] and 63879978.299 [KW], respectively as shown in compost curve figure (6.2). In addition, the grand compost curve (GCC), for the existing plant (Fractionation unit), presented the variation of heat supply and demand as shown in figure (6.3).

6.5 Pinch Technique using Aspen Energy Analyzer

Aspen Energy Analyzer includes a Heat Integration function for analysis and improvement of HEN. It focuses on both the design and operation aspects of the HEN. The operation features of HEN are related to the current operation of a plant, while the design features elucidate the discrepancy between the actual and ideal operation. The Aspen Energy Analyzer can help designers to minimize the gap between the actual and ideal operation. Prior to carrying out heat integration, an input of process constraints and requirements known as the Design is required by the Aspen Energy Analyzer.

6.6 Design (A-Design 1)

The modification which was implemented includes the addition of a new heat exchanger, re-sequence and splitting. This process is automated in the Aspen Energy Analyzer. The software uses a superstructure that includes all feasible structures. The below steps show all of the modifications needed to achieve the best HEN option to minimize the Total Annual Cost (TAC) and increase the energy recovery of the network.

- Two Heat exchangers were added named E-108 and E-109.
- Ten Heaters were added named E-100, E-101, E-106, E-107, E-110, E-116, E-117, E-118, E-119 and E-120
- Four Heaters were deleted by automatic design module named E-100, E-101, E-106 and E-107.
- Ten Coolers were added named E-102, E-103, E-104, E-105, E-111, E-112, E-113, E-114, E-115 and E-121.
- Four Coolers were deleted by automatic design module named E-102, E-103, E-104 and E-105.

- Four splitter-mixers were added by automatic design module named TEE-100-MIX-100, TEE-101-MIX-101, TEE-102-MIX-102 and TEE-103-MIX-103.

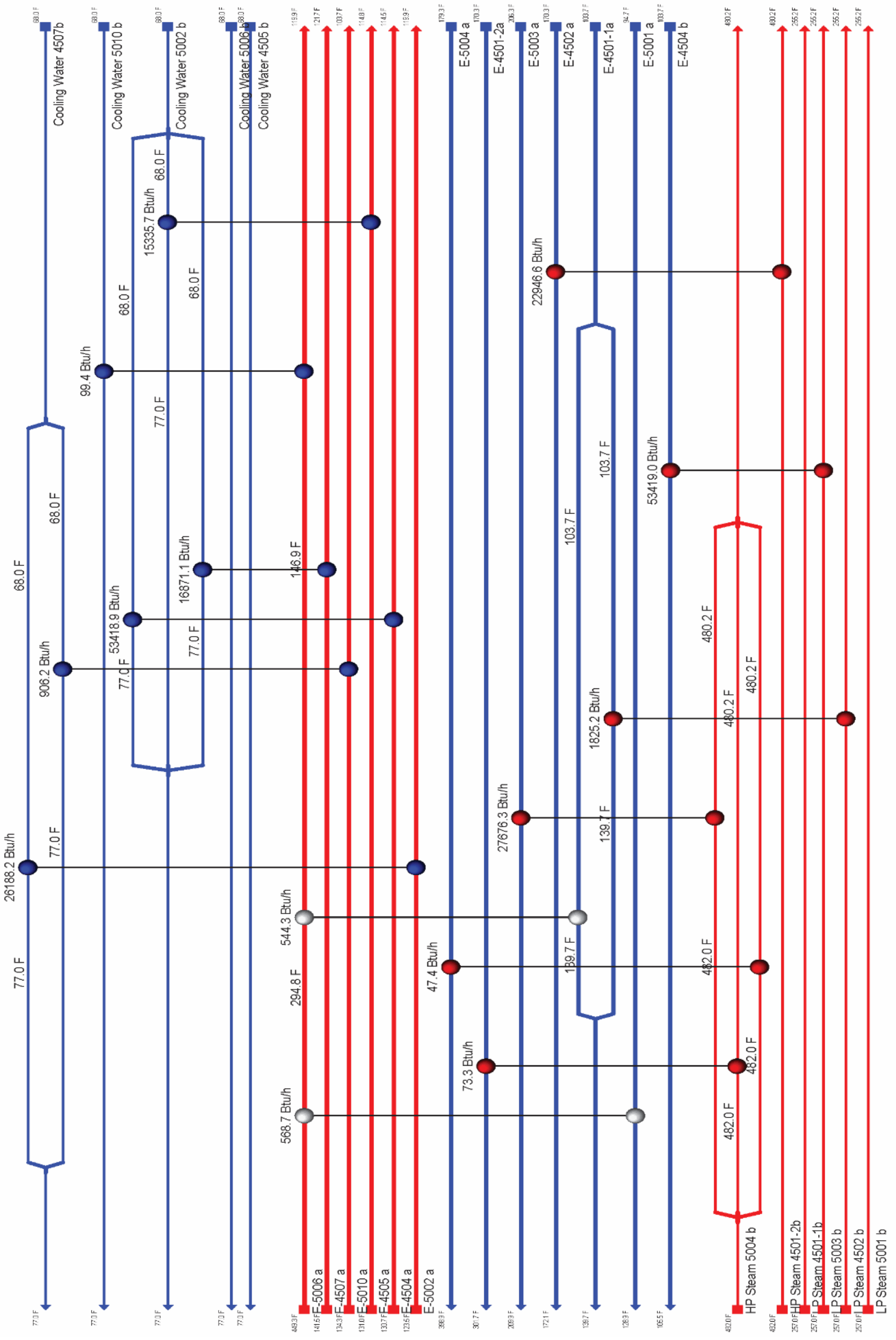

Figure 6.4: The Grid Diagram of design A-Design 1

After analysis, an alternative HEN is planned to save energy. The different HEN is achieved by adding a new two heat exchanger and changing the operating conditions. The Summary and Performance result for the best selection (design A-Design 1), after the above modification, is shown in table (6.6).

Table 6.6: Information about design A-Design 1

Network Cost indexes	
	Cost Index
Heating [MM$/year]	0.214404
cooling [MM$/year]	0.022168
Operating [MM$/year]	0.236573
Capital [MM$]	0.1
Total Cost [MM$/year]	0.417928
Network Performance	
	HEN
Heating [KW]	310619.61108
cooling [KW]	330641.31329
Number of Unit	14.00

Table 6.7: Summary for design A-Design 1

Heat Exchanger	Area [m^2]	load [KW]
E-108	11.9	16.663
E-116	96.13	1565.176
E-118	0.0613	2.149
E-121	0. 462	2.912
E-119	0.0765	1.389
E-109	4.29	15.949
E-111	148.3	767.314
E-117	27.6	810.914
E-120	20.2	672.337
E-110	47.7	53.479
E-112	5.64	26.551
E-114	85.5	449.336
E-113	239.3	1565.175
E-115	80.2	494.324

Table 6.8: Summary of Utilities for final design (A-Design 1)

Utility	Load [KW]
LP Steam 4501-1b	0.00
Cooling Water 4505 b	0.00
Cooling Water 4507b	794.06
LP Steam 5001 b	0.00
Cooling Water 5002 b	2509.44
LP Steam 5003 b	1565.56
HP Steam 5004 b	814.65
Cooling Water 5006 b	0.00
Cooling Water 5010 b	2.91
LP Steam 4502 b	53.49
HP Steam 4501-2b	672.50

Table 6.9: Specification of heat exchanger Network in the design A-Design 1

Heat exchanger	load [KW]	Area [m^2]	Hot stream	Hot T_{in} [^{o} C]	Hot T_{out} [^{o} C]	Cold stream	Cold T_{in} [^{o} C]	Cold T_{out} [^{o} C]
E-108	16.66	11.9	E-5006 a	231.9	146.0	E-5001 a	34.9	53.9
E-116	1565.18	96.2	LP Steam 5003 b	125.0	124.0	E-4504 b	39.8	40.9
E-118	2.15	6.2	HP Steam 5004 b	250.0	249.0	E-4501-2a	76.8	149.8
E-121	2.91	5.0	E-5006 a	63.9	48.9	Cooling Water 5010 b	20.0	25.0
E-119	0.13	76.5	HP Steam 5004 b	250.0	249.0	E-5004 a	81.8	203.9
E-109	15.95	4.3	E-5006 a	146.0	63.9	E-4501-1a	39.9	59.9
E-111	767.31	148.3	E-5002 a	50.9	48.9	Cooling Water 4507b	20.0	25.0
E-117	810.91	27.7	HP Steam 5004 b	250.0	249.0	E-5003 a	96.9	98.9
E-120	672.34	20.2	HP Steam 4501-2b	250.0	249.0	E-4502 a	76.9	77.9
E-110	53.48	4.8	LP Steam 4502 b	125.0	124.0	E-4501-1a	39.9	59.9
E-112	26.55	5.7	E-5010 a	56.9	39.9	Cooling Water 4507b	20.0	25.0
E-114	449.34	85.5	E-4505 a	55.0	46.0	Cooling Water 5002 b	20.0	25.0
E-113	1565.18	239.3	E-4504 a	54.9	45.9	Cooling Water 5002 b	20.0	25.0
E-115	494.32	80.2	E-4507 a	60.9	49.9	Cooling Water 5002 b	20.0	25.0

6.7 Manipulating HI Project Options in Aspen Energy Analyzer

Utility exchanger alteration algorithm are measured simultaneously to achieve more energy saving in the existing HENs and whilst reduce retrofit cost significantly.

Optimize design after network evolution using Aspen Energy Analyzer Optimizer by Utility exchanger modification algorithm method as follows:

1- Select a different method for the program's utility allocation process by choosing the appropriate radio button in the Utility Load Allocation Method group. the Utility Load Allocation method section is:
   - Grand Composite Curve (GCC) Based
   - Supplied Utility Load
   - Cheapest Utility Principle

2- Modify the heat transfer coefficient values by the HTC

3- Modify the utilities by change the Utility Database

4- We select the Use EO Solver. This selected by default. Since, Aspen Energy Analyzer has incorporated a much more robust area optimization algorithm. This new algorithm allows to run constrained area optimization using a novel EO-based model. we put constraints to Maximum Extra Area and Minimum Approach Temperature on each exchanger

5- Specify a value in the following fields:
   - Maximum number of add Exchanger Designs:

   Aspen Energy Analyzer attempts to locate five solutions for this option.

- Maximum number of move two Exchangers Designs:

Aspen Energy Analyzer attempts to locate five solutions for this option.

- Maximum number of move one Exchanger Designs:

Aspen Energy Analyzer attempts to locate five solutions for this option.

- Minimum LMTD Correction Factor

Pinch method for retrofitting Network

At the optimal stage, the potential for energy savings is set by the economy of the process. The best energy recovery corresponds to a minimum approached temperature of 10 ^{o}C, and the lowest energy consumption of the process is given by composite and Grand Composite Curves. After analysis, an alternate HEN is anticipated to save energy by using a utility exchanger modification algorithm. In addition, the following option was generated.

6.7.1 Design Option 1 (A- Design 1-1U).

In this option, a re-piped modification was considered and a retrofit of the HEN was achieved by a clone from A_Design1, where the hot side of Heater E-117 is moved from stream HP Steam 5004 b to stream LP Steam 4501-1b. Also, the area of the heat exchanger [HE117] was increased by adding 130 m^2. This modification led to saving 0.015228 MM$/year, and an investment of 0.87963 [MM$] paying back over a period of 5.7 years. As shown in figure (6.5). Table (6.10) shows the modification results based on retrofitting the base case.

Table 6.10: Summary of result for comparison between the Base case (A-Design1) and Retrofitting network (A- Design1-1U)

	Retrofitting Design (A- Design1-1U)	Base Case (A-Design1)
Heating Cost Index [MM$/year]	0.199176	0.214404
Heating load [KW]	3106.44	3107.09
Cooling Cost Index [MM$/year]	0.022168	0.022168
Cooling load [KW]	3305.61	3305.61
New Area [m^2]	130	3107.09-
New Area Cost [MM$]	0.0879262	-
Operating Savings [MM$/year]	0.015228	-
Payback [years]	5.7	-

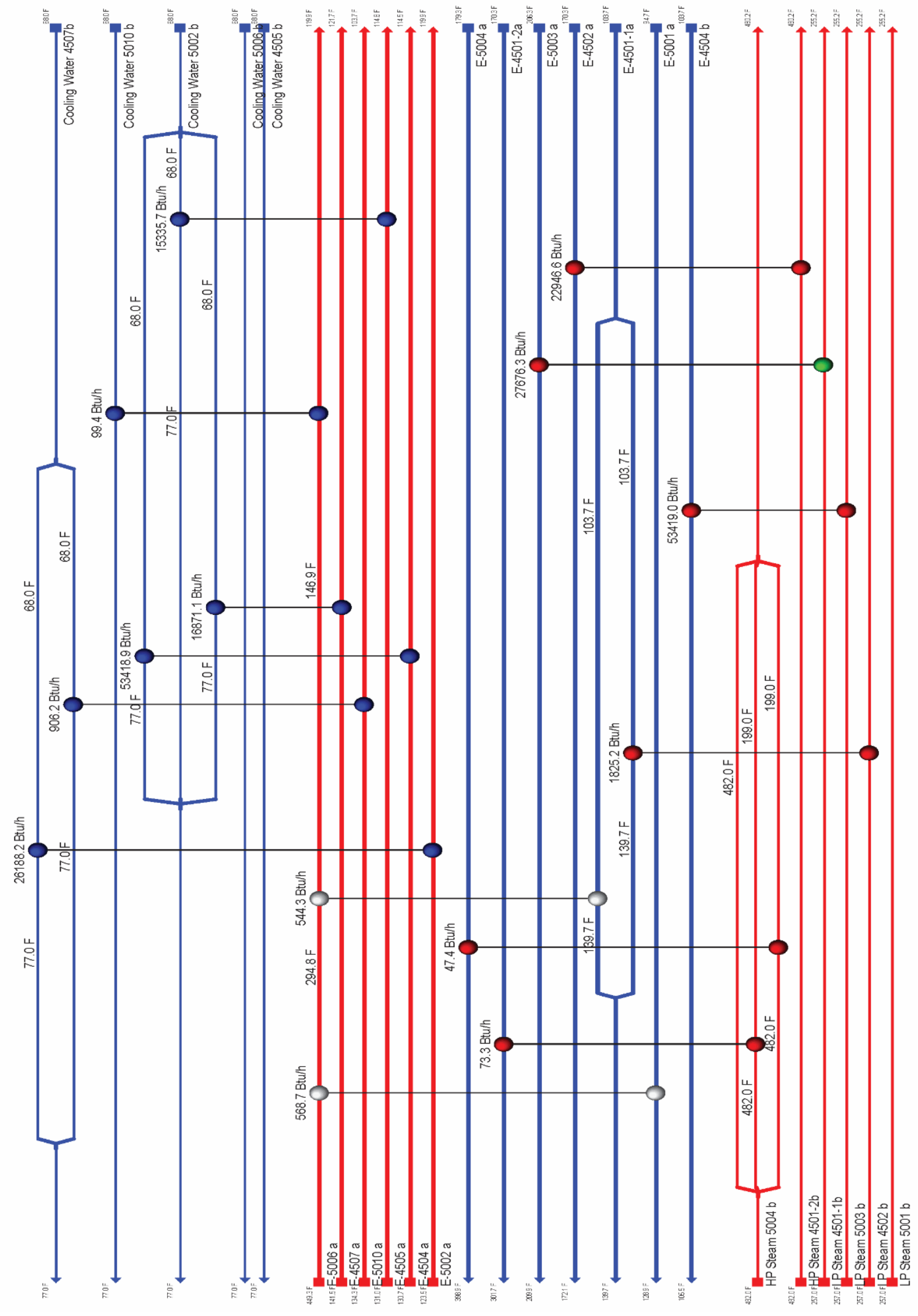

Figure 6.5: Grid Diagram of Design Option 1 (A- Design 1-1U)

Table 6.11: Specification of heat exchanger Network in the final design (A- Design 1-1U).

Heat exchanger	load [KW]	Area [m^2]	Hot stream	Hot T_{out} [^oC]	Hot T_{out} [^oC]	Cold stream	Cold T_{out} [^oC]	Cold T_{out} [^oC]
E-108	16.663	11.9	E-5006 a	231.9	146.0	E-5001 a	34.9	53.9
E-116	1565.176	96.123	LP Steam 5003 b	125.0	124.0	E-4504 b	39.9	40.9
E-118	2.149	0.215	HP Steam 5004 b	250.0	92.8	E-4501-2a	76.9	149.9
E-121	2.912	0.462	E-5006 a	63.9	48.9	Cooling Water 5010 b	20.0	25.0
E-119	1.389	0.368	HP Steam 5004 b	250.0	92.8	E-5004 a	81.9	203.9
E-109	15.949	4.29	E-5006 a	146.0	63.9	E-4501-1a	39.9	59.9
E-111	767.314	148.27	E-5002 a	50.9	48.9	Cooling Water 4507b	20.0	25.0
E-117	810.914	157.34	LP Steam 4501-1b	125.0	124.0	E-5003 a	96.9	98.9
E-120	672.337	20.184	HP Steam 4501-2b	250.0	249.0	E-4502 a	76.9	77.9
E-110	53.479	4.77	LP Steam 4502 b	125.0	124.0	E-4501-1a	39.9	59.9
E-112	26.551	5.64	E-5010 a	56.9	39.9	Cooling Water 4507b	20.0	25.0
E-114	449.336	85.511	E-4505 a	55.0	46.0	Cooling Water 5002 b	20.0	25.0
E-113	1565.175	239.249	E-4504 a	54.9	45.9	Cooling Water 5002 b	20.0	25.0
E-115	494.324	80.1839	E-4507 a	60.9	49.9	Cooling Water 5002 b	20.0	25.0

6.7.2 Design Option 2 (A-Design1- 2U)

In this option, a re-piped modification was considered and a retrofit of the HEN was achieved by cloned from A_Design1-1U, where the hot side of Heater E-118 is moved from stream HP Steam 5004 b to stream HP Steam 4501-2b. Also, the area of the heat exchanger [HE118] was increased by adding 130 m^2. This modification led to saving 0.015051 MM$/year, and an investment of 0.873192 [MM$] paying back over a period of 5.81 years. As shown in figure (6.6). Table (6.12) shows the modification results based on retrofitting the base case.

Table 6.12: The modification results based on retrofitting the base case.

	Retrofitting Design (A- Design1-2U)	Base Case (A-Design1)
Heating Cost Index [MM$/year]	0.199353	0.214404
Heating load [KW]	3108.20	3110.08
Cooling Cost Index [MM$/year]	0.022168	0.022168
Cooling load [KW]	3306.41	3306.41
New Area [m^2]	130	-
New Area Cost [MM$]	0.0873192	-
Operating Savings [MM$/year]	0.0150514	-
Payback [years]	5.81	-

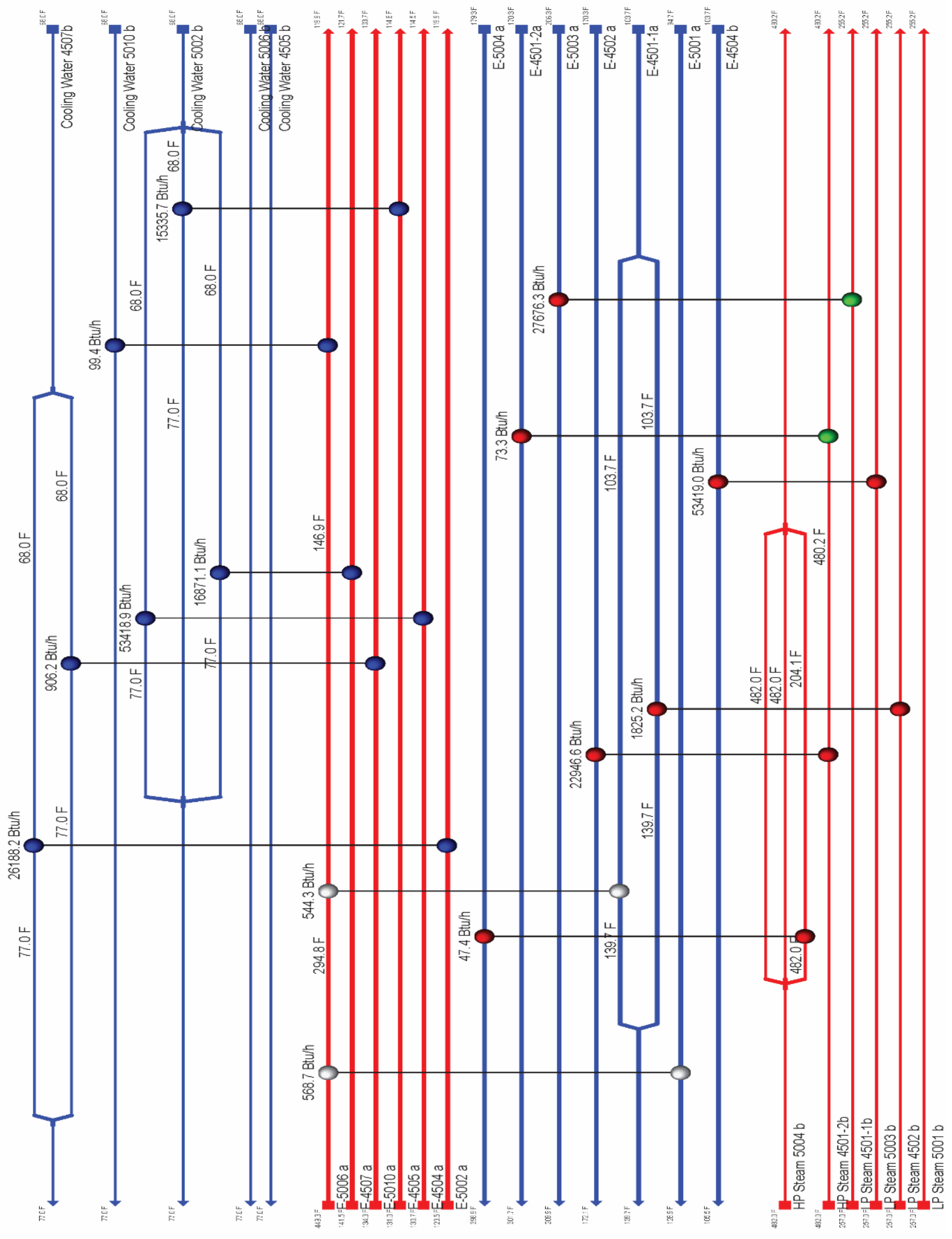

Figure 6.6: Grid Diagram of Design Option 2 (A-Design1- 2U)

6.7.3 Design Option 3 (A-Design1- 3U)

In this option, a re-piped modification was considered and a retrofit of the HEN was achieved by a cloned from A-Design1- 2U, where the hot side of Heater E-119 is moved from stream HP Steam 5004 b to stream HP Steam 4501-2b. Also, the area of the heat exchanger [HE119] was increased adding 130 m^2. This modification led to saving 0.015358 MM$/year, and an investment of 0.863751 [MM$] paying back over a period of 5.63 years. As shown in figure (6.7). Table (6.13) shows the modification results based on retrofitting the base case.

Table 6.13: The modification results based on retrofitting the base case.

	Retrofitting Design (A- Design1-3U)	Base Case (A- Design1)
Heating Cost Index [MM$/year]	0.199046	0.2144.04
Heating load [Btu/ h]	3096.19	3106.19
Cooling Cost Index [MM$/year]	0.022168	0.022168
Cooling load [Btu/ h]	3306.41	3306.41
New Area [m^2]	130	-
New Area Cost [MM$]	0.0863751	-
Operating Savings [MM$/year]	0.015358	-
Payback [years]	5.63	-

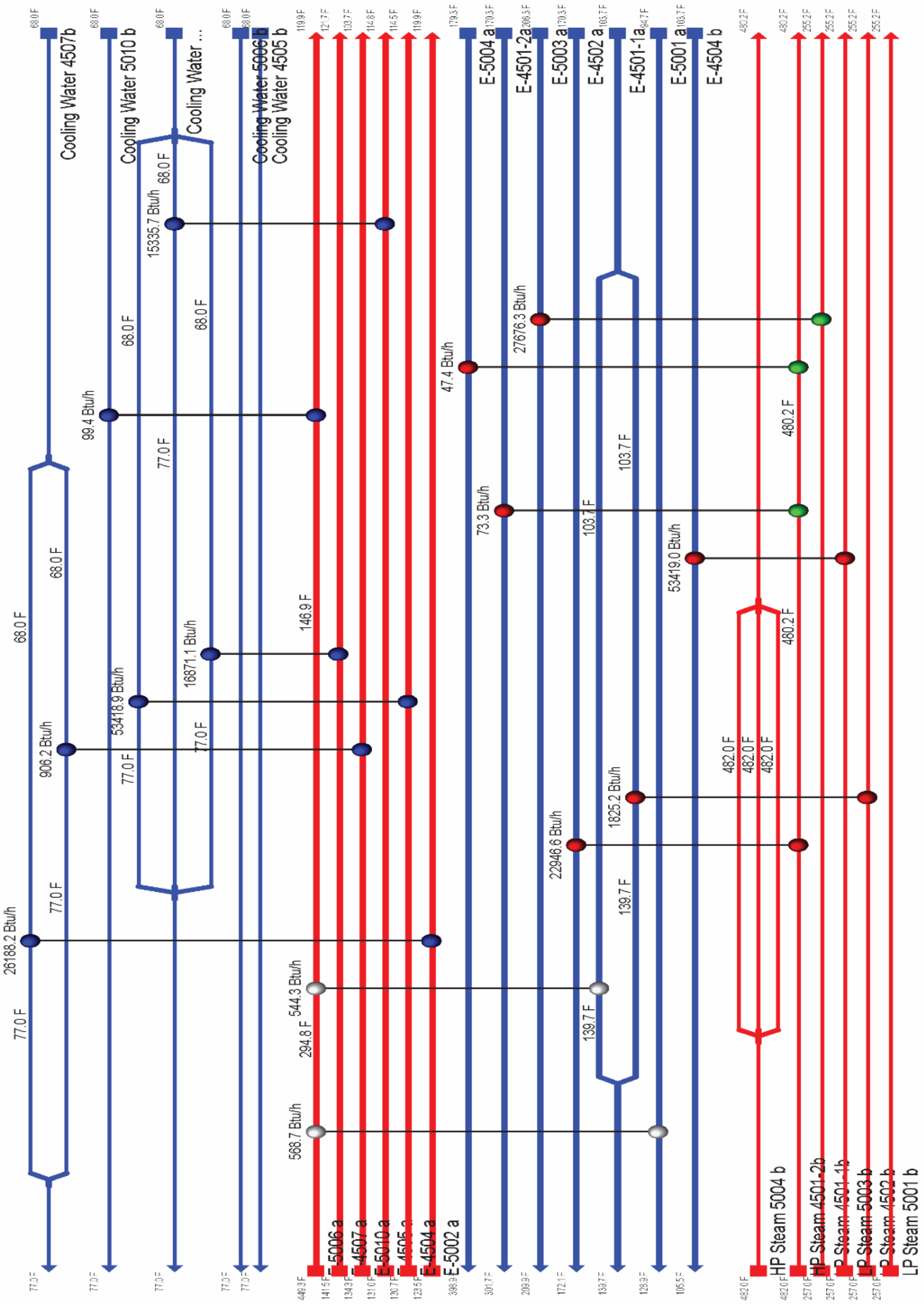

Figure 6.7: Grid Diagram of Design Option 3 (A-Design1- 3U)

6.7.4 Design Option 4 (A- Design1- 4U)

In this option, a re-piped modification was considered and a retrofit of the HEN was achieved by a clone from A-Design1-3U, where the hot side of Heater E-120 is moved from stream HP Steam 4501-2b to stream LP Steam 5001 b. Also, the area of the heat exchanger [HE120] was increased by adding 183 m^2. This modification led to saving 0.028092 MM$/year, and an investment of 0.12681 [MM$] paying back over a period of 4.52 years. As shown in figure (6.8). Table (6.14) shows the modification results based on retrofitting the base case.

Table 6.14: The modification results based on retrofitting the base case.

	Retrofitting Design (A- Design1-4U)	Base Case (A-Design1)
Heating Cost Index [cost/year]	0.186313	0.214404
Heating load [KW]	3106.21	3135.50
Cooling Cost Index [cost/year]	0.022168	0.022168
Cooling load [KW]	3306.41	3306.41
New Area [m^2]	183	-
New Area Cost [MM$]	0.12681	-
Operating Savings [MM$/year]	0.0280915	-
Payback [years]	4.52	-

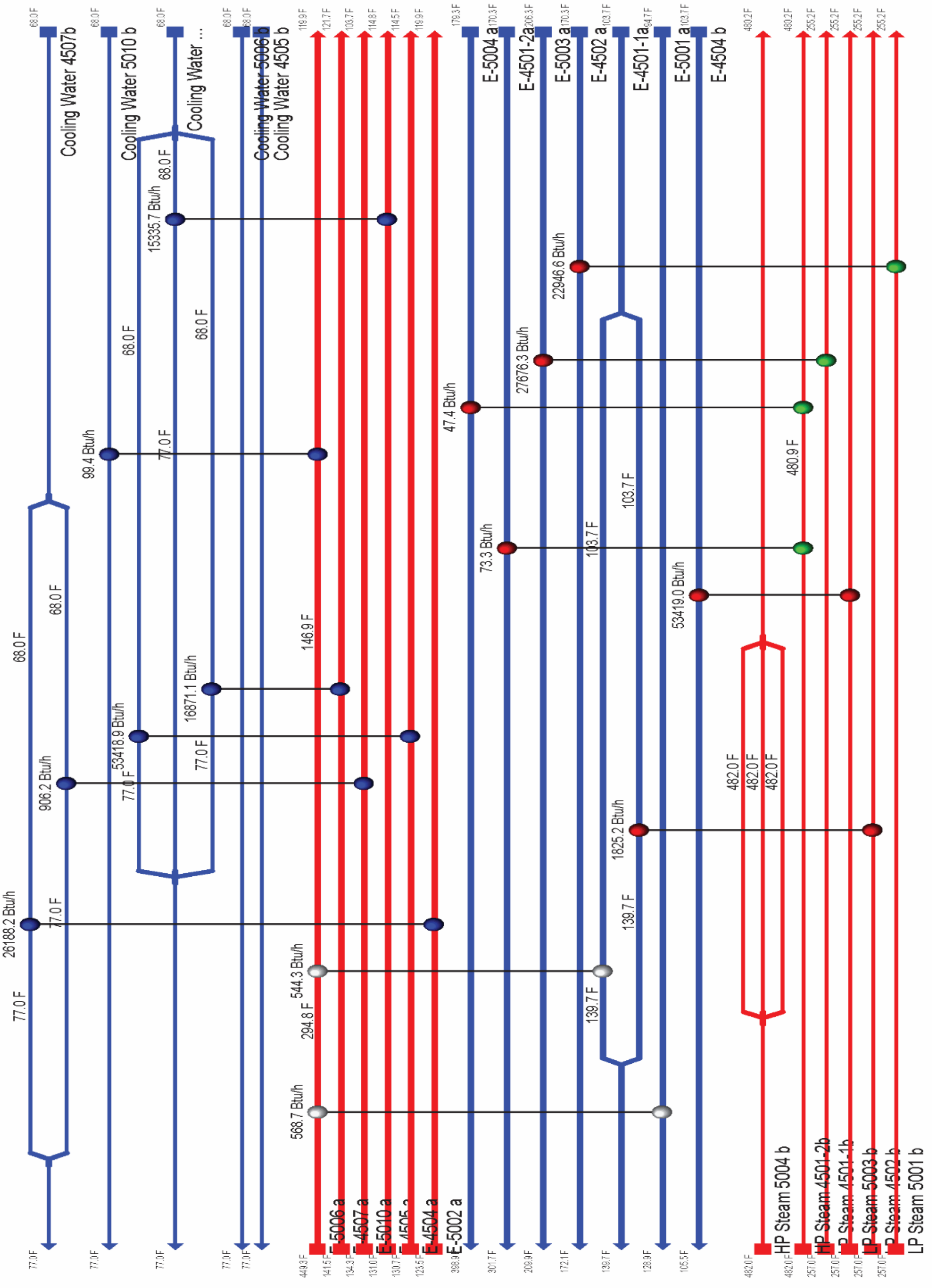

Figure 6.8: Grid Diagram of Design Option 4 (A-Design1- 4U)

6.8 Conclusions

In the existing case study a fractionation unit of HEN was designed by using the Aspen Analyzer software and the pinch technique. Several heat exchangers were designed and, by comparing the TAC, the best design was designated. Also, intensifying heat transfer in a HEN retrofit with a utility exchanger modification algorithm has been successfully addressed in this paper. Compared with the current methods, the planned approach can find valid retrofitted structures for revamping HEN with a heat transfer increase, which can achieve significant energy savings without the high cost of too many modifications. The retrofit design solution was generated by modifying the utility heat exchanger. The generated network options were optimized to trade off capital with energy recovery. The best option design, which was suggested for the revamp, was design option 4 (A-Design1- 4U). This has the less payback period, which is 4.52 years

CHAPTER 7
CONCLUSIONS AND RECOMMENDATIONS

7.1 Conclusions

In this study, energy optimization analysis, targeting and retrofitting of existing complex industrial petrochemical plant were conducted. Stochastic optimization methods were applied to achieve a highly sustainable energy system.

Two techniques were adopted to optimize and retrofit the heat exchanger network (HEN), (1) the Simulated Annealing (SA) and (2) fixed structure technique using MINLP formulation. In the Simulated Annealing technique, the following constraints were specified to search for the optimum solution: (a) the number of the new heat exchanger, (b) the number of re-sequencing, (c) the number of re-piping, (c) the number of splitters. On the other hand, the fixed structure method was applied to get the minimum total annual operating cost based on (a) a minimum change in the structure of HEN, (b) different process stream rate, and (c) varying process stream heat exchanger duties and heat exchanger areas. The results of optimization can be applied to the PDH and fractionation processes by selecting the feasible revamp scheme. The results of our improved HENs were compared to those of existing HEN.

The first proposed design method was energy efficient and was used to find the optimum HEN. A reduction of the Total Annual Cost (TAC) was obtained by optimizing the HEN for a propene dehydrogenation unit (PDH) using the pinch analysis technique. By SA method the TAC of the HEN decreased from 141.07 MM$/y to 120.42 MM$/y. While the fixed structure optimization reduced the TAC from 141.07 MM$/y to 130.00 MM$/y. The results of Simulated Annealing optimization were better than the fixed structure. The suggested new design by SA method has a saving of 20.65

million US $/year, which constitutes saving of 14.64 % as compared to saving of 7.85% by fixed structure optimization.

The design development and improvement method that is an energy efficient technique, was used to find the optimum HEN. A reduction of the Total Annual Cost was obtained by optimizing the HEN for Fractionation unit using the pinch analysis technique. A revamp design was suggested based on the Simulated Annealing (SA) and the fixed structure techniques. The results of SA optimization were better than the fixed structure. The suggested modern design has a saving of 4.17 million US $/year (save up to 20.01%).

A fractionation unit of HEN was designed by using the Aspen Analyzer software V9 and the pinch technique. Compared with the current methods, the planned approach can find valid retrofitted structures for revamping HEN with a heat transfer increase, which can achieve significant energy savings without the high cost of too many modifications. The retrofit design solution was generated by modifying the utility heat exchanger. The generated network options were optimized to trade off capital cost with energy recovery. The best option design, which was suggested for the revamp, was design option 4 (A-Design1- 4U). This has the shortest payback period, which is 4.52 years.

7.2 Recommendations

Based on the current work, some recommendations are made for the successful implementation of the optimization techniques for the process improvement. Further work is needed to improve the HENs for PDH plant and fractionation section, by the development of a standard model of both units in Aspen HYSYS with the aim to reduce the time required for simulation.

In addition to that, some other alternative design of revamping optimization can be applied such as Anew Match, Existing Match, and Reallocation and compared with the current and previous work to show the optimal. There is also need of the employing the realistic data with less assumption to get the real design. Work must be done to include the pressure drop and the fouling rate. Also, several patterns can be tried in the revamping heat exchanger design, several, containing shell- side growth (helical baffles and external fins) and tube –side growth (coiled – wire insert, twisted –tape insert) with the goal to increase the HENs energy recovery within Total Annual Cost.

Furthermore, advanced optimization algorithm such Genetic algorithm (GA), swarm optimization (PSO), and sequential programing (SQP) can be employed for getting better solution.

References

[1] E. M. Al-Mutairi and O. J. Odejobi, “Investigating the thermodynamics and economics of operating the thermal power plant under uncertain conditions,” *Energy Convers. Manag.*, vol. 75, pp. 325–335, Nov. 2013.

[2] M. Escobar and J. O. Trierweiler, “Optimal heat exchanger network synthesis: A case study comparison,” *Appl. Therm. Eng.*, vol. 51, no. 1–2, pp. 801–826, 2013.

[3] S.-G. Yoon, J. Lee, and S. Park, “Heat integration analysis for an industrial ethylbenzene plant using pinch analysis,” *Appl. Therm. Eng.*, vol. 27, no. 5–6, pp. 886–893, Apr. 2007.

[4] R. Smith, M. Jobson, and L. Chen, “Recent development in the retrofit of heat exchanger networks,” *Appl. Therm. Eng.*, vol. 30, no. 16, pp. 2281–2289, 2010.

[5] T. F. Yee and I. E. Grossmann, “Simultaneous optimization models for heat integration - II. Heat exchanger network synthesis,” *Comput. Chem. Eng.*, vol. 14, no. 10, pp. 1165–1184, 1990.

[6] J. M. Zamora and I. E. Grossmann, “A comprehensive global optimization approach for the synthesis of heat exchanger networks with no stream splits,” *Comput. Chem. Eng.*, vol. 21, pp. S65–S70, 1997.

[7] J. Aaltola, “Simultaneous synthesis of flexible heat exchanger network,” *Appl. Therm. Eng.*, vol. 22, no. 8, pp. 907–918, Jun. 2002.

[8] A. J. Isafiade and D. M. Fraser, “Interval based MINLP superstructure synthesis of heat exchanger networks for multi-period operations,” *Chem. Eng. Res. Des.*, vol. 88, no. 10, pp. 1329–1341, Oct. 2010.

[9] T. Gundersen and I. E. Grossmann, “Improved optimization strategies for automated heat exchanger network synthesis through physical insights,” *Comput. Chem. Eng.*, vol. 14, no. 9, pp. 925–944, Sep. 1990.

[10] C. A. Floudas and I. E. Grossmann, “Synthesis of flexible heat exchanger networks with uncertain flowrates and temperatures,” *Comput. Chem. Eng.*, vol. 11, no. 4, pp. 319–336, Jan. 1987.

[11] S. A. Papoulias and I. E. Grossmann, “A structural optimization approach in process synthesis—II: Heat recovery networks,” *Comput. Chem. Eng.*, vol. 7, no. 6, pp. 707–721, 1983.

[12] T. F. Yee and I. E. Grossmann, “Simultaneous optimization models for heat integration—II. Heat exchanger network synthesis,” *Comput. Chem. Eng.*, vol. 14, no. 10, pp. 1165–1184, Oct. 1990.

[13] E. Worrell and C. Galitsky, “Energy efficiency improvement and cost saving opportunities for petroleum refineries,” 2005.

[14] “16. Linnhoff B., and D. R. Vredeveld, Pinch Technology Has - CHE - 4273.” [Online]. Available: https://www.coursehero.com/file/p4qdt4/16-Linnhoff-B-and-D-R-Vredeveld-Pinch-Technology-Has-Come-Of-Age-Chem-Eng/. [Accessed: 09-Mar-2017].

[15] W. V. N. Zhang, “Design of flexible heat exchanger network for multi-period operation,” *Chem. Eng. Sci.*, vol. 61, no. 23, pp. 7730–7753, Dec. 2006.

[16] M. M. El-Halwagi, *Process integration*, vol. 7. Academic Press, 2006.

[17] L. L. Myint and M. M. El-Halwagi, “Process analysis and optimization of biodiesel production from soybean oil,” *Clean Technol. Environ. Policy*, vol. 11, no. 3, pp. 263–276, 2009.

[18] T. Strang and C. Linnhoff-Popien, “A context modeling survey,” in *Workshop Proceedings*, 2004.

[19] E. M. Al-Mutairi, “Retrofit and Exergy Analysis of Fluid Catalytic Cracking Unit Using Heat Recovery Approach Temperature as Decision Criterion,” *Arab. J. Sci. Eng.*, vol. 39, no. 5, pp. 3403–3414, May 2014.

[20] V. Briones and A. C. Kokossis, “Hypertargets: a Conceptual Programming approach for the optimisation of industrial heat exchanger networks — II. Retrofit design,” *Chem. Eng. Sci.*, vol. 54, no. 4, pp. 541–561, Feb. 1999.

[21] U. V. Shenoy, *Heat exchanger network synthesis: process optimization by energy and resource analysis*. Gulf Professional Publishing, 1995.

[22] M. R. Galli and J. Cerdá, “A customized MILP approach to the synthesis of heat recovery networks reaching specified topology targets,” *Ind. Eng. Chem. Res.*, vol. 37, no. 6, pp. 2479–2495, 1998.

[23] J. M. Ponce-Ortega, M. Serna-González, and A. Jiménez-Gutiérrez, “Synthesis of multipass heat exchanger networks using genetic algorithms,” *Comput. Chem. Eng.*, vol. 32, no. 10, pp. 2320–2332, Oct. 2008.

[24] K. F. Huang, E. M. Al-mutairi, and I. A. Karimi, “Heat exchanger network synthesis using a stagewise superstructure with non-isothermal mixing,” *Chem. Eng. Sci.*, vol. 73, pp. 30–43, May 2012.

[25] S. Siddiq, S. Khushnood, Z. U. Koreshi, M. T. Shah, and A. H. Qureshi, “Optimal Energy Recovery from Ammonia Synthesis in a Solar Thermal Power Plant,” *Arab. J. Sci. Eng.*, vol. 38, no. 9, pp. 2569–2577, Sep. 2013.

[26] B. Linnhoff and E. Hindmarsh, “The pinch design method for heat exchanger networks,” *Chem. Eng. Sci.*, vol. 38, no. 5, pp. 745–763, 1983.

[27] B. Linnhoff, D. R. Mason, and I. Wardle, "Understanding heat exchanger networks," *Comput. Chem. Eng.*, vol. 3, no. 1–4, pp. 295–302, Jan. 1979.

[28] D. M. Fraser and N. Hallale, "Retorfit of mass exchange networks using pinch technology - ProQuest," *AIChE J.*, vol. 46, no. 10, p. 2112, 2000.

[29] A. S. Ashaibani and I. M. Mujtaba, "Minimisation of fuel energy wastage by improved heat exchanger network design—an industrial case study," *Asia-Pacific J. Chem. Eng.*, vol. 2, no. 6, pp. 575–584, Nov. 2007.

[30] J. M. Zamora and I. E. Grossmann, "A global MINLP optimization algorithm for the synthesis of heat exchanger networks with no stream splits," *Comput. Chem. Eng.*, vol. 22, no. 3, pp. 367–384, Feb. 1998.

[31] B. Bakhtiari and S. Bedard, "Retrofitting heat exchanger networks using a modified network pinch approach," *Appl. Therm. Eng.*, vol. 51, no. 1–2, pp. 973–979, 2013.

[32] I. E. Grossmann and C. A. Floudas, "Active constraint strategy for flexibility analysis in chemical processes," *Comput. Chem. Eng.*, vol. 11, no. 6, pp. 675–693, Jan. 1987.

[33] A. R. Ciric and C. A. Floudas, "Heat exchanger network synthesis without decomposition," *Comput. Chem. Eng.*, vol. 15, no. 6, pp. 385–396, Jun. 1991.

[34] M. Galli and J. Cerdá, "A customized MILP approach to the synthesis of heat recovery networks reaching specified topology targets," *Ind. Eng. Chem. Res.*, 1998.

[35] L. Myint and M. El-Halwagi, "Process analysis and optimization of biodiesel production from soybean oil," *Clean Technol. Environ. Policy*, 2009.

[36] W. B. Dolan, P. T. Cummings, and M. D. Le Van, “Algorithmic efficiency of simulated annealing for heat exchanger network design,” *Comput. Chem. Eng.*, vol. 14, no. 10, pp. 1039–1050, Oct. 1990.

[37] G. Athier, P. Floquet, L. Pibouleau, and S. Domenech, “Optimization of heat exchanger networks by coupled simulated annealing and NLP procedures,” *Comput. Chem. Eng.*, vol. 20, pp. S13–S18, Jan. 1996.

[38] B. A. Al-Riyami, J. Klemeš, and S. Perry, “Heat integration retrofit analysis of a heat exchanger network of a fluid catalytic cracking plant,” *Appl. Therm. Eng.*, vol. 21, no. 13–14, pp. 1449–1487, 2001.

[39] M. A. Gadalla, “A novel graphical technique for Pinch Analysis applications: Energy Targets and grassroots design,” *Energy Convers. Manag.*, vol. 96, pp. 499–510, 2015.

[40] A. Nemet, J. J. Klemeš, P. S. Varbanov, and V. Mantelli, “Heat Integration retrofit analysis???an oil refinery case study by Retrofit Tracing Grid Diagram,” *Front. Chem. Sci. Eng.*, vol. 9, no. 2, pp. 163–182, 2015.

[41] B. Linnhoff and J. R. Flower, “Synthesis of heat exchanger networks: I. Systematic generation of energy optimal networks,” *AIChE J.*, vol. 24, no. 4, pp. 633–642, Jul. 1978.

[42] R. Smith, *Chemical Process: Design and Integration*, Revised. Wiley, 2005.

[43] B. S. Babaqi, M. S. Takriff, S. K. Kamarudin, N. T. A. Othman, and M. M. Ba-Abbad, “Energy optimization for maximum energy saving with optimal modification in Continuous Catalytic Regeneration Reformer Process,” *Energy*, vol. 120, pp. 774–784, 2017.

[44] F. J.-M. de Meyer, M. Venturoli, and B. Smit, “Molecular simulations of lipid-

mediated protein-protein interactions," *Biophys. J.*, vol. 95, no. 4, pp. 1851–1865, 2008.

[45] E. S. Fraga, "A rewriting grammar for heat exchanger network structure evolution with stream splitting," *Eng. Optim.*, vol. 41, no. 9, pp. 813–831, Sep. 2009.

Vitae

Name	Amir Abdelrazig Merghani Mohammed
Nationality	Sudanese
Date of Birth	7/1/1975
Email	amirmergani@gmail.com
Permanent Address	Faculty of Engineering and technology, Department of Chemical Engineering, University of Gezira
Present Address	King Fahd University of Petroleum and Minerals, 31261, Dhahran, Saudi Arabia
Academic Background	B.Sc. in Chemical Engineering University of Gezira, Wad Medani, Sudan, September 2001. M.Sc. in Chemical Engineering University of Gezira, Wad Medani, Sudan, December 2008. Ph.D. in Chemical Engineering King Fahd University of Petroleum and Minerals, Dhahran, Saudi Arabia. May 2017.